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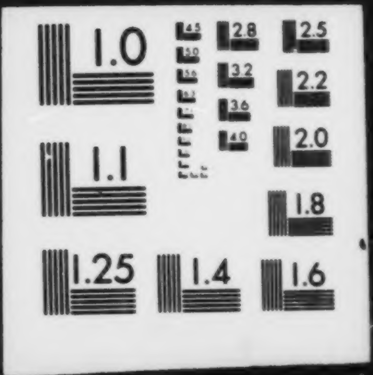
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Process Design Manual for Suspended Solids Removal

Hazen and Sawyer, New York

Prepared for

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FOR
SUSPENDED SOLIDS REMOVAL

U.S. ENVIRONMENTAL PROTECTION AGENCY
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January 1975

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ABSTRACT

This manual surveys current practice in the removal of suspended solids in both traditional and advanced treatment of municipal wastewater. Specific processes are described, design considerations are discussed and results are illustrated by data from actual installations.

Included are processes such as sedimentation, straining and granular media filtration which affect physical separation of solids as well as coagulation and flocculation processes which alter solids characteristics to facilitate such separation. Detailed information is also provided concerning handling and application of coagulant chemicals.

Aspects of operation and maintenance pertinent to design are discussed and estimated costs of construction and operation are provided for particular processes.

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FOREWARD

The formation of the United States Environmental Protection Agency marked a new era of environmental awareness in America. This Agency's goals are national in scope and encompass broad responsibility in the areas of air and water pollution, solid wastes, pesticides, and radiation. A vital part of EPA's national water pollution control effort is the constant development and dissemination of new technology for wastewater treatment.

It is now clear that only the most effective design and operation of wastewater treatment facilities, using the latest avaiable techniques, will be adequate to meet the future water quality objectives and to ensure continued protection of the nation's waters. It is essential that this new technology be incorporated into the contemporary design of waste treatment facilities to achieve maximum benefit of our pollution control expenditures.

The purpose of this manual is to provide the engineering community and related industry a new source of information to be used in the planning, design and operation of present and future wastewater treatment facilities. It is recognized that there are a number of design manuals, manuals of standard practice, and design guidelines currently available in the field that adequately describe and interpret current engineering practices as related to traditional plant design. It is the intent of this manual to supplement this existing body of knowledge by describing new treatment methods, and by discussing the application of new techniques for more effectively removing a broad spectrum of contaminants from wastewater.

Much of the information presented is based on the evaluation and operation of pilot, demonstration and full-scale plants. The design criteria thus generated represent typical values. These values should be used as a guide and should be tempered with sound engineering judgment based on a complete analysis of the specific application.

This manual is one of several available through the Technology Transfer Office of EPA to describe recent technological advances and new information. This particular manual was initially issued in October of 1971 and this edition represents the first revision to the basic text. Future editions will be issued as warranted by advancing state-of-the-art to include new data as it becomes available, and to revise design criteria as additional full-scale operational information is generated.

CHAPTER I

INTRODUCTION

1.1 Purpose

This manual is intended to provide:

1. A basis for selection of processes to meet specific suspended solids (SS) removal requirements
2. A basis for design of particular processes
3. A basis for selection of particular equipment configurations for a given process.

Since the emphasis is on information applicable to design or modifications of solids removal facilities, only those processes are included for which reliable data from actual applications are available.

1.2 Wastewater Solids

The *total solids* in wastewater exist in a distribution of sizes from individual ions up to visible particles. Specific analytical procedures (1) have been established to distinguish the suspended fraction of the total solids and to further distinguish the settleable fraction within the SS. A typical concentration of SS for raw domestic wastewaters is 200 mg/l, but this can vary substantially from system to system (see below). The lower limiting size for the SS fraction (about 1.5 microns) is arbitrarily defined by the test procedures and it should be noted that variations in test procedures themselves can also lead to widely varying results, especially at the low solids levels characteristic of treated effluents.

Other workers (2) (3) (4) have applied procedures which distinguish four solids fractions, and determine proportions of other wastewater characteristics such as COD, Nitrogen, Volatile (organic) matter in each fraction. For a New Jersey municipal raw wastewater, solids distribution in terms of these fractions was found to be as follows (2):

		Raw Wastewater		Secondary Effluent	
Fraction	Size Range Microns	Total Solids mg/l	Volatile Matter mg/l	Total Solids mg/l	Volatile Matter mg/l
Soluble	<0.001	351	116	312	62
Colloidal	0.001-1	31	23	8	6
Supra-Colloidal	1-100	57	43	28	24
Settleable	>100	74	59	0	0

The settleable and supracolloidal fractions together are essentially equivalent to the suspended fraction referred to above. Dividing lines between fractions again are somewhat arbitrary depending on tests applied, and overall concentrations in different fractions can vary substantially between systems depending on factors such as water use, travel time in sewers, ground-water infiltration, and prevalence of home garbage grinding. Contributions of dissolved, colloidal and suspended solids from individual homes, multi-family dwellings or other point sources often have concentrations two or more times the average for a whole system (5).

In addition to particle size, specific gravity and strength or shear resistance of wastewater solids may affect solids separation performance. The three basic types of solids separation processes—gravity separation, physical straining, and granular media filtration are discussed in Chapters 7, 8 and 9, respectively. Wastewater solids characteristics can be altered to enhance performance of the above separation processes. Chapters 4 and 6 discuss chemical treatment (precipitation and/or coagulation) and physical treatment (flocculation) aimed at alteration of solids characteristics. In addition, during the separation processes themselves, agglomeration and compaction of solids generally continues, increasing separation efficiency and reducing the volume of separated solids.

Biological wastewater treatment processes also affect solids characteristics and hence solids separation. Activated sludge solids have been found (6) to have a distinct bimodal distribution with one mode in the supracolloidal to settleable range and another near the border between the colloidal and supracolloidal fractions. The concentrations and size limits in each range are affected by conditions in the biological reactor (Chapter 6). Dean (7) has noted that bacteria, cellular debris, etc. fall into the finer (colloidal-supracolloidal) range. Agglomeration of these finer solids generally increases the efficiency of subsequent separation processes.

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CHAPTER 2

GENERAL DESIGN CONSIDERATIONS

2.1 Applications of Suspended Solids Separation Processes

Processes for SS separation may fill three distinct functions in wastewater treatment.

1. Pretreatment to protect subsequent processes and reduce their loadings to required levels
2. Treatment to reduce effluent concentrations to required standards
3. Separation of solids to produce concentrated recycle streams required to maintain other processes.

In the first two functions effluent quality is the prime consideration, but where the third function must be fulfilled along with one of the others, design attention must be given to conditions for both the separated solids (sludge) and the process effluent.

Table 2-1 compares several SS separation process applications selected to illustrate how their performance and their loading requirements are functions of their applications.

Wedge-wire screens can operate at very high hydraulic and solids loadings, but do not greatly reduce SS. Hence, wedge wire screens are limited to pretreatment applications where subsequent processes will assure production of a satisfactory final effluent. They can be considered as an adjunct to primary sedimentation or, where conditions prescribe, as an alternative.

Sedimentation units must operate at relatively low hydraulic loadings (large space requirements), but can accept high solids loadings. With proper chemical or biological pretreatment and design, they can produce good quality effluents.

Microscreens and granular-media filters, operating at significantly higher hydraulic loads than sedimentation units, can produce an effluent with lower SS than is possible with sedimentation alone. In general they are not designed to accept high solids loadings, and are normally used following other processes which put out relatively low effluent SS concentrations.

2.2 Process Selection

Selection of one of the alternative processes can be based on cost only where all factors not reflected in cost are equivalent. Direct cost comparison of individual solids removal processes usually proves impossible because of differences in factors such as: 1) effluent quality, 2) pretreatment requirements, 3) effects on sludge processing, 4) housing, space and head requirements.

TABLE 2-1

SELECTED SS SEPARATION PROCESS APPLICATIONS

Type of Separation Process	Application	Typical Loading Ranges: Hydraulic gpm/sq ft	Typical Loading Ranges: Infl. Solids mg/l	Typical Loading Ranges: Infl. Solids lb/day/sq ft	Expected Effluent SS$^{(a)\,(b)}$ mg/l	Remarks
Straining						
Wedge Wire Screens	Preliminary Treatment of Raw Wastewater	10-30	200	25-75	150-190	
Microscreens	Polishing of Biological System Effluent	3-10	30	1-2	5-15	
Gravity Separation						
Plain Sedimentation	Primary Treatment	0.4-1.6	200	0.5-2	120-80	
Chemical Coagulation and Settling	Chemical Treatment of Raw Sewage (Phosphate Removal Levels)	0.3-1.0	200$^{(c)}$	1-6	20-60	Chemical treatment for 90% + phosphorus removal.
Plain Sedimentation (secondary)	Separation of Solids after Activated Sludge Treatment	0.25-0.75	2000-5000	4-40	10-50	Upper effluent quality Limit may increase with poor biological treatment. Allowable solids loadings depends on solids characteristics.
Granular Media Filtration	Polishing of Biological Effluent or Filtration of Chemically-Coagulated and	4-8	30	1-2	5-15	
	Settled Raw Wastewater or	3-5	40	1-2	10-20	
	Secondary Effluent	3-5	5-10	1-2	1-3	Secondary treatment may be biological or by activated carbon.

(a) Based on raw wastewater SS of 200 mg/l.

(b) Performance is highly dependent on character of solids applied and hence on conditions in prior treatment.

(c) Influent solids do not include chemical solids.

Meaningful cost comparisons usually involve practically the entire process configuration of the treatment facility, including processes for disposal of solid residues, and reflect how the individual unit processes affect one another.

Cost data on individual processes for suspended solids removal are given in Chapter 10. Outlined below are some additional factors not reflected in the unit process cost figures, but which may warrant consideration in overall comparisons.

1. Sludge Handling. Where chemical treatment is used to remove BOD or phosphates or improve SS removals, significant quantities of chemical sludge are produced. The cost of disposal of this sludge must be considered in process selection unless configurations being compared involve similar chemical treatment. The actual cost involved will depend greatly on the particular method of sludge disposal to be used. For general guidance, in a 10 mgd plant using thickening, digestion, vacuum filtration and landfill for sludge disposal, chemical addition of 200 mg/l alum would increase the sludge disposal costs by almost 20 percent (from 9.7 cents/1000 gal to 11.5 cents/1000 gal of plant flow on a 1972 cost basis) (Chapter 10, Reference 14).

   Where this difference appears significant in the comparison of alternatives for SS removal, specific sludge disposal figures should be included in process comparisons. Information on expected sludge quantities from particular chemical treatment processes is provided in Chapter 4.

2. Buildings. The need for housing specific unit process varies with climate and other local conditions. Where the housing requirements of alternative processes obviously differ widely under particular local conditions, building cost should be considered in the selection.

3. Land Requirements. Generally, land requirements are a small enough factor in overall cost that the differences for various process alternatives are not significant. Where adequate land is unavailable or very costly, however, area requirements of alternative processes should be compared in detail. Minimum land requirements may be estimated at between 1.20 (large plants) and 2.0 (small plants) times the area of the process units themselves.

4. Head requirements. Some of the processes employed for SS separation (sedimentation, microscreens, etc.) require relatively small head (only 2 to 3 ft. to overcome losses at inlet and effluent controls and in connecting piping). Others, such as granular-media filters, and wedge-wire screens, require greater differential head (10 ft or more). Differences in head requirements are most significant where they necessitate capital outlay for an extra pumping step. The costs for pumping, however, even with lifts above 10 ft. are usually not large in relation to the overall costs for treatment facilities.

CHAPTER 3

FLOW VARIATIONS AND EQUALIZATION

3.1 Flow Variation

Both the rate and characteristics of the inflow to most treatment plants vary significantly with time. Diurnal cycles are found in all domestic discharges. Weekly and seasonal cycles are common in municipal systems as are variations between wet and dry weather.

Even where only domestic flows are involved, the magnitude of variations can differ widely between different systems depending on system configuration, water use habits of the population and opportunities for groundwater infiltration or direct inflow of surface or subsurface drainage. Industrial and institutional flows where significant, can further alter domestic patterns.

Because of these wide differences, design of treatment facilities should be based, whenever possible, on measurements of actual flow variations in existing systems. In projects being submitted for federal construction grants, analysis of existing flows is required in any case to identify "excessive" infiltration inflow. Flows are considered excessive if they can be eliminated more cheaply than they can be treated. Projected flow variations from existing systems should reflect elimination of excessive flows.

Flows tend to be less variable in larger systems, due chiefly to differing times of travel from different sections and to damping effects of flow storage in large sewers. Widely varying relations have been reported between peak-to-average or minimum-to-average flow ratios and system size (i.e. average flow or tributary population) (1) (2) (3) (4). Care should be taken in using any of these relations for estimating flow variations in new systems or system additions. In terms of the factors which affect flow variations, applications should be limited to systems similar to those for which the relation was originally developed. Relations for which the basis is unclear should be disregarded.

3.2 Performance vs. Flow Variation

Variations in influent flow rate and characteristics affect performance of all suspended solids removal processes to some degree. Relations between performance and hydraulic or solids loadings are discussed for individual processes in succeeding chapters. Magnitude and character of significant recycled flows resulting from specific processes are also indicated.

Relations between performance and loadings are frequently developed in pilot units run under steady flow conditions, or from data from actual plants compiled without close attention to short-term peaks. In using such relations for design decisions, care must be taken to allow for the effects of short term flow variations on performance. Short term would include any time span less than that for which performance requirements are stated. Typically requirements are on a monthly average basis, often with a less stringent requirement for the worst week or worst day within the month.

Designs based on maximum 24-hour flow, with allowance for diurnal peaks, provide some margin so that weekly or monthly requirements can be met even when other factors affecting process performance are not optimum.

3.3 Flow Equalization

Equalization storage can be used to reduce diurnal variations in flow and in concentration of SS or other wastewater characteristics. Storage may also be used to handle peaks caused by direct inflow to the sewers during wet weather. Assuming that equivalent performance can be obtained either by increasing the size of treatment facilities or by providing equalizing basins, selection between these approaches can be based on their relative costs and environmental impacts. In plants using processes involving large, short-term recycle flows—such as for backwashing granular media filters—equalization is almost always justified.

The EPA Process Design Manual for Upgrading Existing Wastewater Treatment Plants provides a basis for design of equalization facilities to achieve any given degree of equalization of either peak flows or peak flows and solid loadings (5). Material from the Design Manual relevant to flow equalization only is also available in a separate publication (6).

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CHAPTER 4

PRINCIPLES OF CHEMICAL TREATMENT

4.1 Introduction

Chemical coagulation and flocculation are accomplished by a combination of physical and chemical processes which thoroughly mix the chemicals with the wastewater and promote the aggregation of wastewater solids into particles large enough to be separated by sedimentation, flotation, media filtration or straining. The strength of the aggregated particles determines their limiting size and their resistance to shear in subsequent processes.

For particles in the colloidal and fine supracolloidal size ranges (<1 to 2 microns) natural stabilizing forces (electrostatic repulsion, physical separation by absorbed water layers) predominate over the natural aggregating forces (van der Waals) and the natural mechanism (Brownian movement) which tends to cause particle contact. Coagulation of these fine particles involves both destabilization and physical processes which disperse coagulants and increase the opportunities for particle contact. Destabilization, the action of chemical coagulants, is discussed in this chapter. Physical processes, including chemical mixing, flocculation, and solids contact processes, are discussed in Chapter 6.

Chemical coagulants used in wastewater treatment are generally the same as those used in potable water treatment and include: alum, ferric chloride, ferric sulfate, ferrous chloride, ferrous sulfate and lime. The effectiveness of a particular coagulant varies in different applications, and in a given application each coagulant has both an optimum concentration and an optimum pH range.

In addition to coagulants themselves, certain chemicals may be applied for pH or alkalinity adjustment (lime, soda ash) or as flocculating agents (organic polymers). For full effectiveness chemical coagulation requires initial *rapid mixing* (Chapter 6) to thoroughly disperse the applied chemicals so that they can react with suspended and colloidal solids uniformly.

4.2 Destabilization Mechanisms

The destabilizing action of chemical coagulants in wastewater may involve any of the following mechanisms:

1. Electrostatic charge reduction by adsorption of counter ions
2. Inter-particle bridging by adsorption of specific chemical groups in polymer chains
3. Physical enmeshment of fine solids in gelatinous hydrolysis products of the coagulants.

The significance of these mechanisms in design is considered briefly below. Extensive discussion of the mechanisms can be found in the literature (1) (2) (3) (4).

4.2.1 Electrostatic Charge Reduction

Finely dispersed wastewater solids generally have a negative charge. Adsorption of cations from metal salt coagulants (in the case of iron and aluminum from their hydrolysis products), or from cationic polymers can reduce or reverse this charge.

Where electrostatic charge reduction is a significant destabilization mechanism, care must be taken not to overdose with coagulant. This can cause complete charge reversal with restabilization of the oppositely charged coagulant-colloid complex.

4.2.2 Interparticle Bridging

When polymeric coagulants contain specific chemical groups which can interact with sites on the surfaces of colloid particles, the polymer may adsorb to and serve as a bridge between the particles. Coagulation using polyelectrolytes of the same charge as the colloids or non-ionic polymers depends on this mechanism. Restabilization may occur if excessive dosages of polymer are used. In this case all sites on the colloids may adsorb polymer molecules without any bridging. Excessive mixing can also cause restabilization by fracture or displacement of polymer chains.

4.2.3 Enmeshment in Preciptated Hydrolysis Products

Hydroxides of iron, aluminum or, at high pH, magnesium form gelatinous hydrolysis products which are extremely effective in enmeshing fine particles of other material. These hydroxides are formed by reaction of metal salt coagulants with hydroxyl ions from the natural alkalinity in the water or from added alkaline chemicals such as lime or soda ash. Sufficient natural magnesium is frequently present in wastewaters so that effective coagulation is obtained merely by raising the pH with lime. Organic polymers do not form hydrolysis products of significance in this mechanism. At a pH value lower than that required to precipitate magnesium, the precipitates produced by lime treatment are frequently ineffective in enmeshing the colloidal matter in wastewater. The remedy for this condition generally involved addition of low dosage of iron salts or polymers as coagulant aids both to destabilize and to increase the probability of enmeshment of colloids.

Coagulants may also react with other constituents of the wastewater, particularly anions such as phosphate and sulfate, forming hydroloysis products containing various mixtures of ions. The chemistry of the reactions is extremely complex and highly dependent on pH and alkalinity. The presence of high concentrations of these anions may require increased doses of coagulants or pH adjustment to achieve effective removals of SS.

4.3 Selection of Chemical Coagulants

Design of chemical treatment facilities for SS removal must take into account: 1) the types and quantities of chemicals to be applied as coagulants, coagulant aids and for pH control and 2) the associated requirements for chemical handling and feeding (Chapter 5) and for mixing and flocculation after chemical addition (Chapter 6). Reactions of specific coagulant chemicals are detailed in Chapter 5.

Selection of coagulants should be based on jar testing of the actual wastewater (Section 4.5) to determine dosages and effectiveness, and on consideration of the cost and availability of different coagulants. Where expected changes in waste characteristics or market conditions may favor different coagulants at different times, chemical feed and handling should be set up to permit a switchover. In developing a testing program general information on experience at other locations and on costs should be considered to aid in selection of processes and coagulants to be tested.

Experience to date with improved SS removal from chemical coagulation has been almost solely in systems designed to remove phosphorus. Guidelines for design and coagulant selection for such systems are available in another manual (5). Descriptive data and SS removal performance for several existing phosphorus removal installations are summarized in Table 4-1.

Few cases have been reported involving chemical coagulation aimed at SS removal alone without phosphorous removal requirements. Anionic polymers have been used to increase SS removal in primary treatment at Rocky River, Ohio (6). Doses of 0.3 mg/l reduced SS from 107 mg/l to 65 mg/l. Mogelnicki (7) reported use of anionic polymer at a dosage of 1 mg/l to improve primary clarifier SS removal from 43 percent to 76 percent.

In discussing the favorable results sometimes obtained in polymer applications O'Melia (4) warns that it can be a time-consuming task to find the specific conditions (pH, ionic strength, polymer type, molecular weight, degree of hydrolysis, etc.) which will provide economy and effectiveness.

Pilot work at Denver, Colorado (8) on coagulation of effluent from an activated sludge nitrification system showed substantial reductions in SS, turbidity, BOD and other pollution parameters, using lime and alum doses well below those needed for effective phosphate removal. Lime dosages of 100 mg/l were sufficient to reduce SS to below 15 mg/l after settling and 5 mg/l after filtration. Phosphate reduction was less than 80 percent. Alum dosages of about 50 mg/l were sufficient to reduce suspended solids and phosphorus concentrations to similar levels. In direct filtration of alum-coagulated nitrified effluent, SS were reduced to less than 2 mg/l with an alum dosage of 60 mg/l. Phosphate reductions at this alum concentration were only about 65 percent (6-7 mg/l residual). This latter practice is accompanied by shorter filter runs due to significant increases in solids loading.

Calgon Corporation investigated the use of ferric chloride with polymer addition for a small municipal wastewater treatment plant at Leetsdale, Pennsylvania (9). Ferric chloride dosages were less than those necessary for 80 percent phosphate removal. Dosages and SS reductions are shown below:

TABLE 4-1

SS REMOVAL PERFORMANCE FOR CHEMICAL COAGULATION APPLICATIONS TO PHOSPHATE REMOVAL

LOCATION	PROCESS	PLANT SIZE mgd	AVERAGE CHEMICAL FEED mg/l	pH	BASIC EQUIPMENT	SUSPENDED SOLIDS		DATA PERIOD Months	OVERFLOW RATE gpd/sq ft	COMMENTS
						Inf. mg/l	Settled Eff. mg/l			
Lebanon, Ohio	IPC	0.1	Lime ~250	9.5	1-stage SC	109	30	45	1440	Acid pH adjustment. Reference 19
EPA, Blue Plains Plant, Washington, D.C.	IPC	0.1	Lime 460(a) +$FeCl_3$ 5(b)	11.5(a) 10.0(b)	2-stage SC(a) FM,FL,S(b)	158	14.4	6	500-1800	Two stages; with intermediate recarbonation. Reference 20
Ely, Minn.	Tertiary	1.5	Lime 250-350(a) +Polymer .2(a) +$FeCl_3$ 6(b)	11.8(a) 9.5(b)	2-stage SC	75	10	5	570(a) 660(b)	System designed for nearly complete (eff. ≤ 0.05 mg/l) phosphorus removal. Reference 21
S. Lake Tahoe, California	Tertiary	7.5	Lime 400 275	11.3 10.5	FM,FL, S SR	38 38	10 25	21	400-600	Recarbonation Reference 22
Lebanon, Ohio	Tertiary	0.1	Lime 220-270	9.5	SC	43.5	16.5	10	1440	Acid pH adjustment made. Reference 23
Nassau County, New York	Tertiary	0.6	Alum 20	-	SC	22.5	2.5	72	860	Bulking 2-3 times a year. Reference 24
Salt Lake City, Utah	IPC	0.04-0.1	Iron 34-41 Polymer 0-1.5	-	SC	90	21.9	2.5	360-1080	Reference 11
		0.05-0.09	Alum 14 Polymer 0-0.25	-	SC	95	26.9	1	500-870	---
		0.03-0.18	Lime 270-586	9.8-11.0	SC	101	11.6	5	290-1800	H_2SO_4; pH adjustment.
Leetsdale, Pa.		0.6	$FeCl_3$ 100 +Polymer 0.5	6.5-6.7	S	280	38	-	-	Primary Treatment, Constant Feed - Reference 9

Key:
FM - Flash Mix Unit
FL - Flocculator
SC - Solids Contact (Sludge Blanket)
SR - Solids Recirculation
S - Settling
IPC - Independent Physical Chemical

Notes:
Lime as $Ca(OH)_2$
Alum as Al^{+++}
Iron Salts as Fe^{+++}
(a) First Stage
(b) Second Stage
Overflow rates are minimum and maximum where range appears.

Dosage		SS		
FeCl3 mg/l	Polymer mg/l	Influent mg/l	Effluent mg/l	Reduction Percent
74	0.5	160	23	86
50	0.6	71	27	62
37	0.08	85	33	61

Aluminum or iron salts tend to react with soluble phosphate preferentially so that substantial phosphorus removal must be involved before organic colloids can be destabilized (10). Required dosages will be affected by phosphorus content. Similarly lime treatment to a pH at which coagulation is effective precipitates substantial phosphorus. Because chemical dosage and pH range for optimum SS removal may differ somewhat from those for optimum phosphorus removal, coagulant requirements may be determined by the effluent criteria for either pollutant, depending on wastewater characteristics and the choice of chemical.

4.3.1 Sludge Production

Chemical coagulation increases sludge production in sedimentation units due both to greater removal of influent suspended solids and to insoluble reaction products of the coagulation itself. For phosphorus removal, data on sludge production and sludge characteristics and sample procedures for estimating sludge quantities are presented elsewhere (5).

The weight of sludge solids can be estimated by calculation of the sum of the expected SS removal and of the precipitation products expected from the coagulant dosages applied. Usually jar tests can be employed to obtain the necessary information for this calculation.

4.3.2 pH Control and Alkalinity

The critical factor in the control of lime reactions is pH. The pH for optimum effectiveness of lime coagulation, determined from jar testing and process operating experience, can be used as a set point for a pH control of lime dosing.

Alum and iron salt coagulation are much less sensitive to pH. Testing can determine optimum dosages for coagulation and whether natural alkalinity is adequate for the reactions (see Ch. 5). If supplemental alkalinity is needed either regularly or on an intermittent basis (e.g. during high wet weather flows) provisions should be included for feeding necessary amounts of lime or soda ash.

4.3.3 Points of Chemical Addition

In independent physical-chemical treatment or in phosphate removal in the primary clarifier ahead of biological treatment, chemicals are added to raw sewage. In tertiary treatment for phosphate removal and SS reduction, they are added to secondary effluent. In both cases, proper mixing and flocculation units are needed. For phosphate removal or improvement of SS capture in biological secondary treatment, chemicals are often added directly to aeration units or prior to secondary settling units, without separate mixing and flocculation. In some phosphate removal applications coagulants have been added at multiple points, e.g. prior to primary settling and as part of a secondary or tertiary treatment step.

4.3.4 Supplementary Coagulants

Addition of the hydrolyzing metal coagulants to wastewater often results in a small slow-settling floc or precipitate of phosphorus. Additional treatment is required to produce a water with low residual suspended solids. Polymeric coagulants have proved to be quite beneficial in aggregating the precipitation products to a settleable size and increasing the shear strength of the floc against hydraulic breakup (11). Data on particular applications appear in Table 4-1.

4.4 Coagulation Control

Because coagulation represents a group of complex reactions, laboratory experimentation is essential to establish and maintain the optimum coagulant dosage and to determine the effects of important variables on the quality of coagulation of the wastewater under investigation. With alum and iron coagulants two procedures are generally followed for this purpose: the jar test and measurement of zeta potential. Proper control of lime coagulation may be maintained by measuring the pH or automatically titrating alkalinity after lime addition.

4.4.1 Jar Test

The single, most widely used test to determine coagulant dosage and other parameters is the jar test. The equipment for this test and the directions for its proper performance have been published (12) (13) (14) (15). The jar test attempts to simulate the full scale coagulation-flocculation process and has remained the most common control test in the laboratory since its introduction in 1918. Since the intent is to simulate an individual plant's conditions, it is not surprising that procedures may vary but generally have certain common elements. The jar test apparatus consists of a series of sample containers, usually six, the contents of which can be stirred by individual mechanically-operated stirrers. Wastewater to be treated is placed in the containers and treatment chemicals are added while the contents are being stirred. The range of conditions, for example, coagulant dosages and pH, are selected to bracket the anticipated optima. After a short, 1 to 5 minute period of rapid stirring to ensure complete dispersion of coagulant, the stirring rate is decreased and flocculation is allowed to continue for a variable period, 10 to 20 minutes or more, depending on the simulation. The stirring is then stopped and the flocs are allowed to settle for a selected time. The supernatant is then analyzed for the desired parameters. With wastewater the usual analyses are for turbidity or suspended solids, pH, residual phosphorus and residual coagulant.

If desired, a number of supernatant samples may be taken at intervals during the settling period to permit construction of a set of settling curves which provide more information on the settling characteristics of floc than a single sample taken after a fixed settling period. A dynamic settling test may also be used in which the paddles are operated at 2 to 5 rpm during the settling period. This type of operation more closely represents settling conditions in a large horizontal basin with continuous flow.

It should be noted that simple jar tests cannot simulate the conditions in solids contact reactors (Chapter 6) and may indicate somewhat higher coagulant dosages than are actually necessary when using these units for coagulation.

Several six-position stirrers are available commercially for running jar tests; one from Phipps and Bird, (Phipps and Bird, Inc., Richmond, Va.), another from Coffman Industries, (Coffman Industries, Inc., Kansas City, Ka.), are shown in Fig. 4-1. Standard laboratory mixers have also been used; however, it is difficult to obtain reproducible mixing conditions using different pieces of equipment. Various types of containers, usually beakers or jars, are used to hold the samples. Improved mixing may be obtained by adding stationary plates in the containers as described by Camp and Conklin (15). The Coffman stirrer has an attachment which makes it possible to add coagulant to all containers simultaneously. Good results, however, can be obtained by rapidly adding coagulant from a large graduated pipette to each jar in sequence.

A simple apparatus, shown in Fig. 4-2, can be constructed from tubing, rubber stoppers and small aquarium valves to permit rapid sampling of supernatant. The unit is placed next to the sample jars at the beginning of the settling period with the curved stainless steel tubes dipping into the jars. At desired intervals the vent valve is covered with a finger, permitting vacuum to draw samples into the small sample bottles. The needle valves are adjusted so that supernatant is drawn into all the bottles at the same rate. When sufficient sample is obtained, the vent is uncovered and the bottles are replaced with empties. The maximum sampling rate is about once per minute.

Fig. 4-3 shows characteristic types of settling curves which may be obtained. Curve A indicates a coagulation which produced a uniformly fine floc so small that at the end of 1 to 2 minutes settling, the supernatant had a turbidity equal to that of the starting water due, in part, to the fine floc which resisted settling. Settling was slow and the final turbidity was not satisfactory. Curve B represents the most common type of settling rate obtained. During the first 5 minutes, the settling rate was practically a straight line on a semilog plot. Settling was rapid and clarification was satisfactory. The coagulation represented by curve C shows that a mixture of large rapid settling floc and small, slow-settling particles was obtained. Settling was rapid for the first two minutes, but with little further clarification after that. High residual turbidity may also have resulted from incomplete coagulation. Curve D represents the ultimate in coagulation. Practically all of the floc particles were so large and dense that 97 percent settled within three minutes. Sedimentation was essentially complete within that time since only 0.5 percent additional floc settled in the next 27 minutes. Final

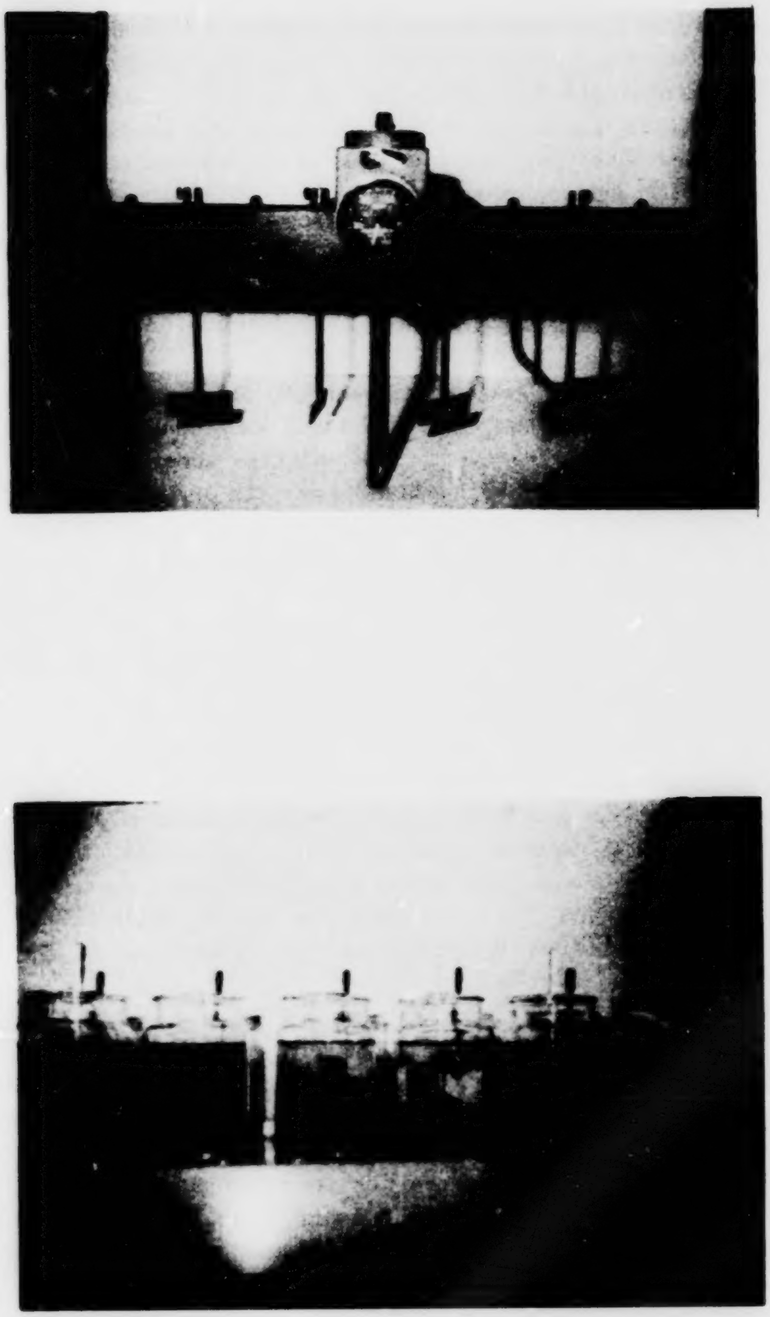


FIGURE 4-1
JAR TEST UNITS WITH MECHANICAL (TOP)
AND MAGNETIC (BOTTOM) STIRRERS

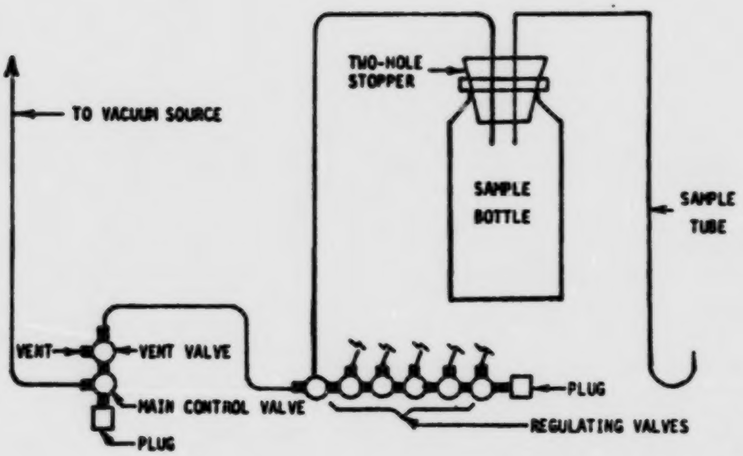
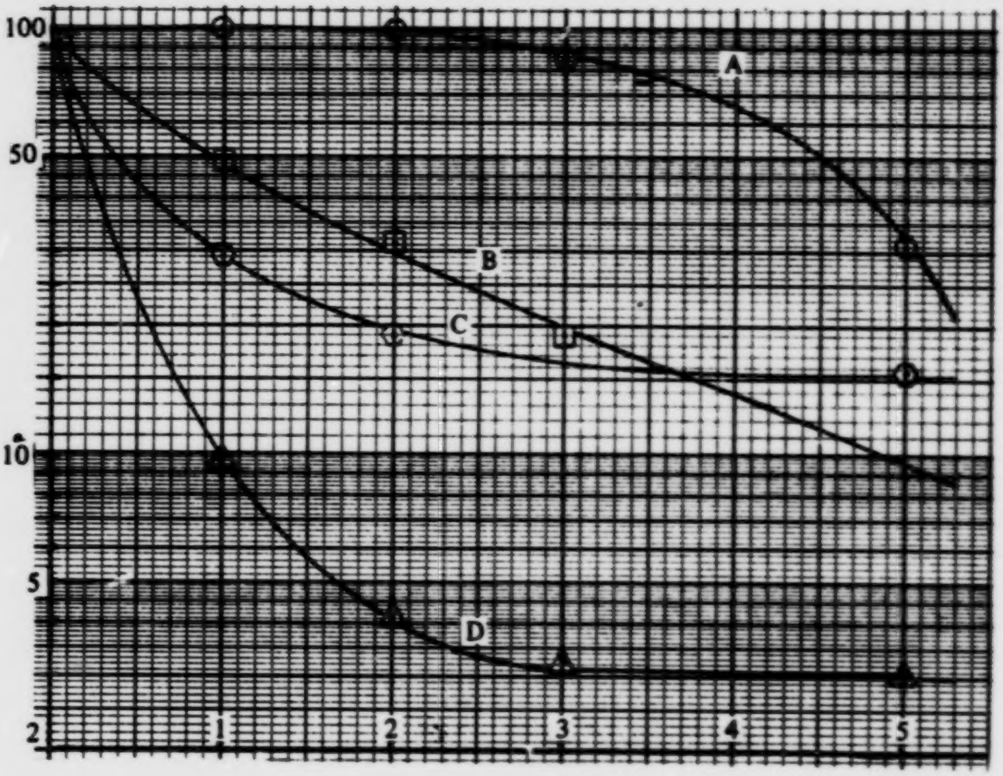

TWO-HOLE STOPPER
TO VACUUM SOURCE
SAMPLE BOTTLE
SAMPLE TUBE
VENT
VENT VALVE
MAIN CONTROL VALVE
PLUG
PLUG
REGULATING VALVES

FIGURE 4-2
SIX-POSITION SAMPLER



SETTLING TIME-MIN

FIGURE 4-3

SETTLING CURVES FREQUENTLY OBTAINED

clarity of the supernatant was entirely satisfactory.

Measurement of turbidity provides the most rapid indication of the degree of solids removal obtained. The recommended procedure for turbidity measurement by light scattering is given in the 13th edition of *Standard Methods for Examination of Water and Wastewater;* however, other methods varying from simple visual evaluation to measurement of light transmitted on a laboratory spectrophotometer can be used for purposes of comparison. Measurement of residual suspended solids is the only procedure which gives the actual weight concentration of solids remaining, but the procedure is too slow for purposes of process control. Where the character of the solids does not vary widely, their concentration generally correlates well with measured turbidity.

A typical jar test might be run as follows:

Wastewater samples are placed in containers and rapid mix is started at 100 rpm. Selected dosages of coagulant covering the expected range of the optimum concentration are rapidly added to the containers and mixed for approximately 1 minute. If a polymer is to be used as a coagulant aid, it is usually added to each jar at or just before the end of the rapid mix. The paddles are then slowed to 30 rpm and mixing continues for 20 minutes. The paddles are then stopped and the sampling apparatus previously-described is placed in position. At settling times of 1,3,5,10 and possibly 20 minutes samples of supernatant are drawn for turbidity measurement. After the final turbidity sample is drawn, a larger volume of supernatant may be decanted for more complete analysis. Results are plotted as in Fig. 4-4 for judgment as to the desired coagulant dosage.

If additional alkalinity is required to hold the coagulation in the optimum pH range, this should be added to the samples ahead of the coagulant unless automatic titrators are set up for pH control.

Once an approximate optimum coagulant concentration has been determined, it may be desirable to repeat the jar test using that optimum with varying quantities of added alkalinity to give different pH values. Experience in coagulating a given wastewater provides the best guide as to methods for controlling the process.

4.4.2 Zeta Potential

Measurement of particle charge is another procedure which may be useful for control of the coagulation process (16) (17) (18). The total particle charge is distributed over two concentric layers of water surrounding the particle: an inner layer of water and ions which is tightly bound to the particle and moves with it through the solution, and an outer layer which is a part of the bulk water phase and moves independently of the particle. Charges of these layers are not directly measureable, but the zeta potential, which is the residual charge at the interface between the layer of bound water and the mobile water phase, can be determined indirectly with commercially-available instruments.

In the zeta potential measurement procedure, a sample of treated water containing floc is placed in a special plastic cell under a microscope as shown in Fig. 4-5. Under the in-

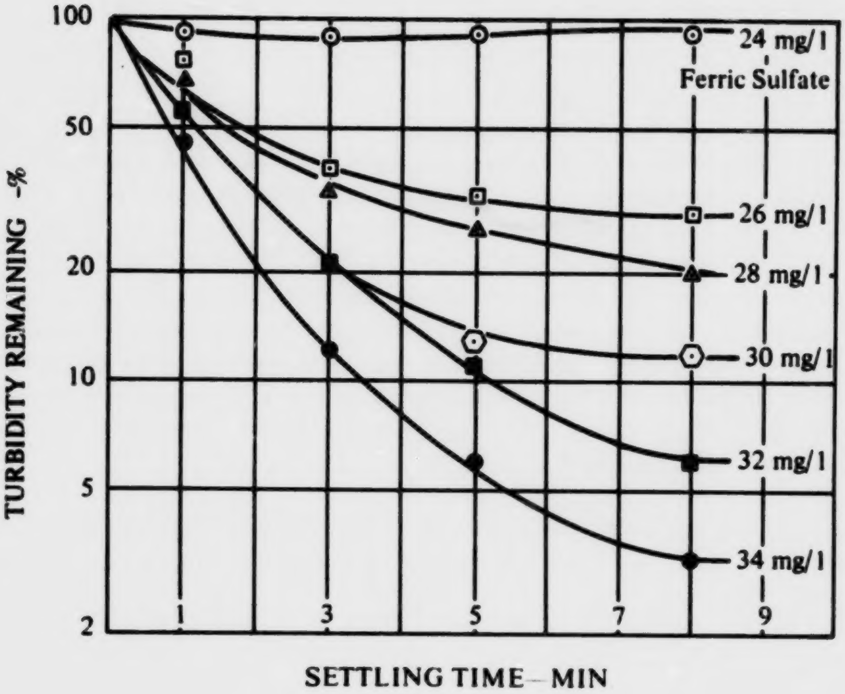


FIGURE 4-4
JAR TEST RESULTS

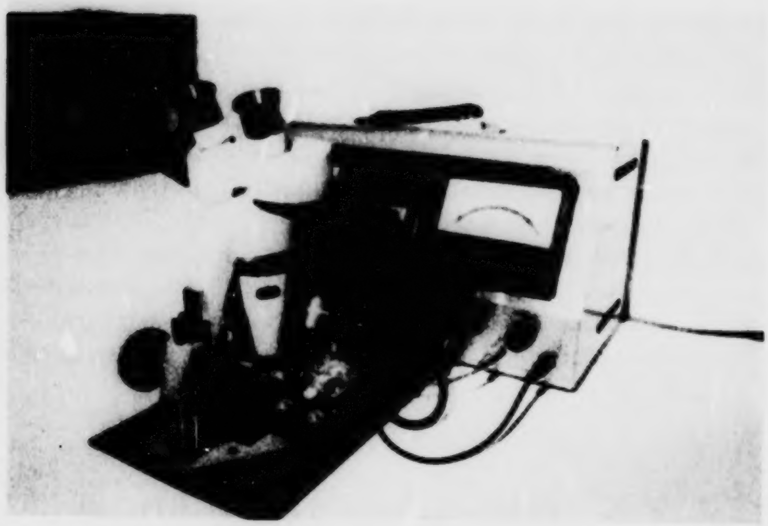


FIGURE 4-5
ZETA POTENTIAL APPARATUS

fluence of a voltage applied to electrodes at the ends of the cell, the charged particles will migrate to the electrode having a polarity opposite that of the particle. The velocity of migration will be proportional to the particle charge and to the applied voltage. The particle velocity can be calculated by observing the time it takes a particle to travel a given distance across an ocular micrometer. The zeta potential can then be obtained from a chart which combines the particle velocity with instrumental parameters. Detailed operating instructions are supplied with the instruments. Because of uncertainties in the constants relating charge and particle mobility, many test results are reported directly in terms of particle mobility.

To control the coagulation by zeta potential, samples of water while being mixed are dosed with different concentrations of coagulant. Zeta potentials are then measured and recorded for floc in each sample. The dosage which produces the desired zeta potential value is applied to the treatment plant. Zeta potentials of floc produced in the plant may also be measured as a means of control. The zeta potential value for optimum coagulation must be determined for a given wastewater by actual correlation with jar tests or with plant performance as in Fig. 4-6. The control point is generally in the range of 0 to 10 millivolts. If good correlations can be obtained between some zeta potential values and optimum plant performance, then it is possible to make rapid measurements of particle charge to compensate for major variations in wastewater composition due to storm flows or other causes. Short term variations such as those due to sudden industrial waste dumps are still beyond control with any present techniques because of the time lag between recognition of a problem with coagulation and adoption of a satisfactory change of coagulation conditions.

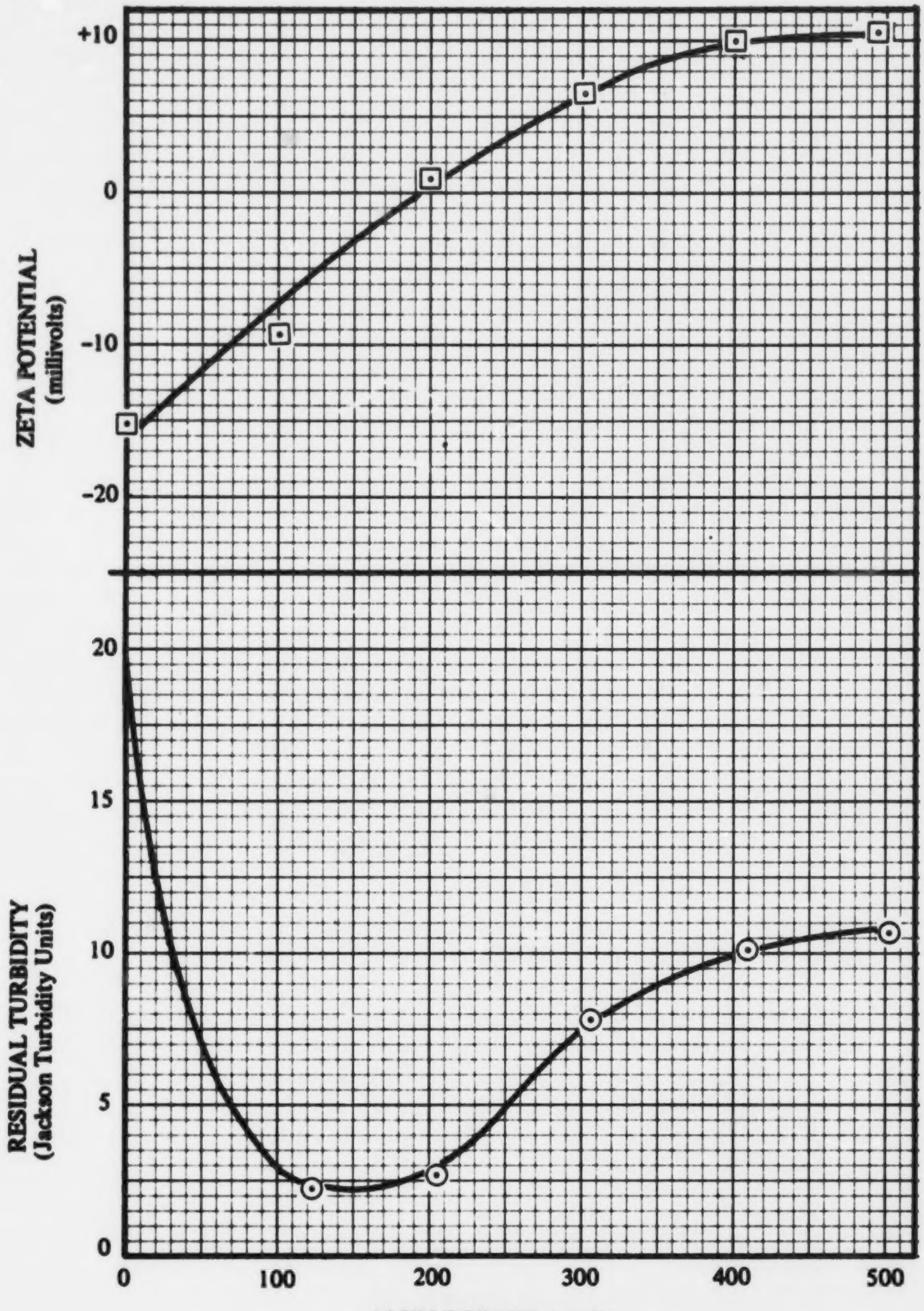


FIGURE 4-6

COAGULATION OF RAW SEWAGE WITH ALUM

4.5 References

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CHAPTER 5

STORAGE AND FEEDING OF CHEMICALS

5.1 General

This chapter surveys the chemicals most commonly used for suspended solids removal, with respect to their properties, availability, storage, transport, reactions and feeding. All chemical costs quoted in this chapter were obtained from the latest issues of "Chemical Marketing Reporter" (Schnell Publishing Co., Inc., New York, N. Y.) available during preparation of this manual. Wide ranges in bagging costs primarily reflect bag sizes that may be ordered. All chemical costs presented are for guidance only and are subject to significant variations due to time and current market conditions. Actually costs for the chemicals being considered should be carefully checked prior to selection.

5.2 Aluminum Compounds

The principal aluminum compounds that are commercially available and suitable for suspended solids removal are dry and liquid alum. Sodium aluminate has been used in activated sludge plants, but for phosphorus removal, and its applicability for suspended solids removal is limited.

5.2.1 Dry Alum

5.2.1.1 Properties and Availability

The commercial dry alum most often used in wastewater treatment is known as "filter alum." and has the approximate chemical formula $Al_2(SO_4)_3 \cdot 14H_2O$ and a molecular weight of about 600. Alum, is white to cream in color and a 1 percent solution has a pH of about 3.5. The commercially available grades of alum and their corresponding bulk densities and angles of repose are:

GRADE	ANGLE OF REPOSE	BULK DENSITY lb./cubic feet
Lump	------------	62 to 68
Ground	43	60 to 71
Rice	38	57 to 71
Powdered	65	38 to 45

Each of these grades has a minimum alumium content of 27 percent, expressed as Al_2O_3, and maximum Fe_2O_3 and soluble contents of 0.75 and 0.5 percent, respectively. Visosity and solution crystallation temperatures are included in the subsequent section on liquid alum.

Since dry alum is only partially hydrated, it is slightly hygroscopic. However, it is relatively stable when stored under the extremes of temperature and humidity encountered in the United States.

The solubility of commercial dry alum at various temperatures is as follows:

Temperature deg F	Solubility lb/gal
32	6.03
50	6.56
68	7.28
86	8.45
104	10.16

Dry alum is not corrosive unless it absorbs moisture from the air, such as during prolonged exposure to humid atmospheres. Therefore, precautions should be taken to ensure that the storage space is free of moisture.

Alum is shipped in 100 lb bags, drums, or in bulk (minimum of 40,000 lb) by truck or rail. Bag shipments may be ordered on wood pallets if desired. Locations of the major production plants are listed in Table 5-1. At present, the price range for dry alum in bulk quantities is $58 to $63/ton. F.O.B. the point of manufacture. Freight costs to the point of usage must be added to this. Bagging adds approximately $4 to 5/ton to the cost.

TABLE 5-1

PARTIAL LIST OF ALUM MANUFACTURING PLANTS

Location	Manufacturer	Form of Alum Available
ALABAMA		
Coosa Pines	American Cyanamid	Liquid
Demopolis	American Cyanamid	Liquid
Mobile	American Cyanamid	Liquid and Dry
Naheola	Stauffer	Liquid
ARKANSAS		
Pine Bluff	Allied	Liquid
CALIFORNIA		
Bay Point (San Francisco)	Allied	Liquid and Dry
El Segundo (Los Angeles)	Allied	Liquid
Richmond (San Francisco)	Stauffer	Liquid
Vernon (Los Angeles)	Stauffer	Liquid
COLORADO		
Denver	Allied	Liquid and Dry

TABLE 5-1 (continued)

Location	Manufacturer	Form of Alum Available
FLORIDA		
Fernandina Beach	Tennessee Corp.	Liquid
Jacksonville	Allied	Liquid
Port St. Joe	Allied	Liquid
GEORGIA		
Atlanta (2 plants)	Burris, Allied	Liquid and Dry
Augusta	Tennessee Corp.	Liquid and Dry
Cedar Springs	Tennessee Corp.	Liquid
Macon	Allied	Liquid
Savannah	Allied	Liquid
ILLINOIS		
E. St. Louis	Allied	Liquid and Dry
Joliet	American Cyanamid	Liquid and Dry
LOUISIANA		
Bastrop	Stauffer	Liquid
Baton Rouge	Stauffer	Liquid
Monroe	Allied	Liquid
New Orleans	Allied	Liquid and Dry
Springhill	Stauffer	Liquid
MAINE		
Searsport	Northern	Liquid and Dry
MARYLAND		
Baltimore	Olin	Dry
MASSACHUSETTS		
Adams	Holland	Liquid
Salem	Hamblet & Hayes	Liquid and Dry
MICHIGAN		
Detroit	Allied	Liquid
Escanaba	American Cyanamid	Liquid
Kalamazoo (2 plants)	Allied, American Cyanamid	Liquid
MINNESOTA		
Cloquet	American Cyanamid	Liquid
Pine Bend	North Star	Liquid and Dry

TABLE 5-1 (continued)

Location	Manufacturer	Form of Alum Available
MISSISSIPPI		
Monticello	American Cyanamid	Liquid
Vicksburg	Allied	Liquid
NEW JERSEY		
Newark	Essex	Liquid and Dry
Warners	American Cyanamid	Liquid and Dry
NORTH CAROLINA		
Acme	Wright	Liquid
Plymouth	American Cyanamid	Liquid
OHIO		
Chillicothe	Allied	Liquid
Cleveland	Allied	Liquid and Dry
Hamilton	American Cyanamid	Liquid and Dry
Middletown	Allied	Liquid
OREGON		
North Portland	Stauffer	Liquid and Dry
PENNSYLVANIA		
Johnsonburg	Allied	Liquid
Marcus Hook	Allied	Liquid and Dry
Newell	Allied	Liquid
SOUTH CAROLINA		
Catawba	Burris	Liquid
Georgetown	American Cyanamid	Liquid and Dry
TENNESSEE		
Chattanooga	American Cyanamid	Liquid and Dry
Counce	Stauffer	Liquid
Springfield	Burris	Liquid
TEXAS		
Houston (2 plants)	Stauffer, Ethyl	Liquid and Dry

TABLE 5-1 (continued)

VIRGINIA		
Covington	Allied	Liquid
Hopewell	Allied	Liquid
Norfolk	Howerton Gowen	Liquid
WASHINGTON		
Kennewick	Allied	Liquid
Tacoma (2 plants)	Stauffer, Allied	Liquid
Vancouver	Allied	Liquid and Dry
WISCONSIN		
Menasha	Allied	Liquid
Wisconsin Rapids	Allied	Liquid

Manufacturers and Addresses

Allied Chemical Corporation
Industrial Chemicals Division
P.O. Box 1139R
Morristown, New Jersey 07960

American Cyanamid Company
Ind. Chem. Div.
P.O. Box 66189
Chicago, Illinois 60666

Burris Chemical Company
Charleston, South Carolina

Essex Chemical Corporation
1402 Broad Street
Clifton, New Jersey 07015

Ethyl Corporation
Houston, Texas

Hamblet & Hayes Company
Colonial Road
Salem, Massachusetts 01970

Holland Chemical Company
Adams, Massachusetts 01220

Howerton Gowen Company, Inc.
Norfolk, Virginia

Northern Chemical Industries, Inc.
Searsport, Maine 04974

North Star Chemicals Inc.
P.O. Box 28-T
South St. Paul, Minnesota

Olin Corporation,
Chemicals Division
745 Fifth Avenue
New, York, New York 10022

Stauffer Chemical Company
299 Park Avenue
New York, New York 10017

Tennessee Corporation
Cities Service Company
Industrial Chemicals Division
P.O. Box 50360
Atlanta, Georgia

Wright Chemical Co.
Acme, North Carolina

5.2.1.2 General Design Considerations

Ground and rice alum are the grades most commonly used by utilities because of their superior flow characteristics. These grades have less tendency to lump or arch in storage and therefore provide more consistent feeding qualities. Hopper agitation is seldom required with these grades, and in fact may be detrimental to feeding because of the possibility of packing the bin.

Alum dust is present in the ground grade and will cause minor irritation of the eyes and nose on breathing. A respirator may be worn for protection against alum dust. Gloves may be worn to protect the hands. Because of minor irritation in handling and the possibility of alum dust causing rusting of adjacent machinery, dust removal equipment is desirable. Alum dust should be thoroughly flushed from the eyes immediately and washed from the skin with water.

5.2.1.3 Storage

A typical storage and feeding system for dry alum is shown in Figure 5-1. Bulk alum can be stored in mild steel or concrete bins with dust collector vents located in, above, or adjacent to the equipment room. Recommended storage capacity is about 30 days. Dry alum in bulk can be transferred with screw conveyors, pneumatic conveyors, or bucket elevators made of mild steel. Pneumatic conveyor elbows should have a reinforced backing as the alum can contain abrasive impurities.

Bags and drums of alum should be stored in a dry location to avoid caking. Bag or drum loaded hoppers should have a nominal storage capacity for eight hours at the nominal maximum feed rate so that personnel are not required to charge the hopper more than once per shift. Converging hopper sections should have a minimum slope of 60 degree to prevent arching.

Bulk storage hoppers should terminate at a bin gate so that the feeding equipment may be isolated for servicing. The bin gate should be followed by a flexible connection, and transition hopper chute or hopper which acts as a conditioning chamber over the feeder.

5.2.1.4 Feeding Equipment

The feed system includes all of the components required for the proper preparation of the chemical solution. Capacities and assemblies should be selected to fulfill individual system requirements. Three basic types of chemical feed equipment are used: volumetric, belt gravimetric, and loss-in-weight gravimetric. Volumetric feeders are usually used where initial low cost and usually lower capacities are the basis of selection. Volumetric feeder mechanisms are usually exposed to the corrosive dissolving chamber vapors which can cause corrosion of discharge areas. Manufacturers usually control this problem by use of an electric heater to keep the feeder housing dry or by using plastic components in the exposed areas.

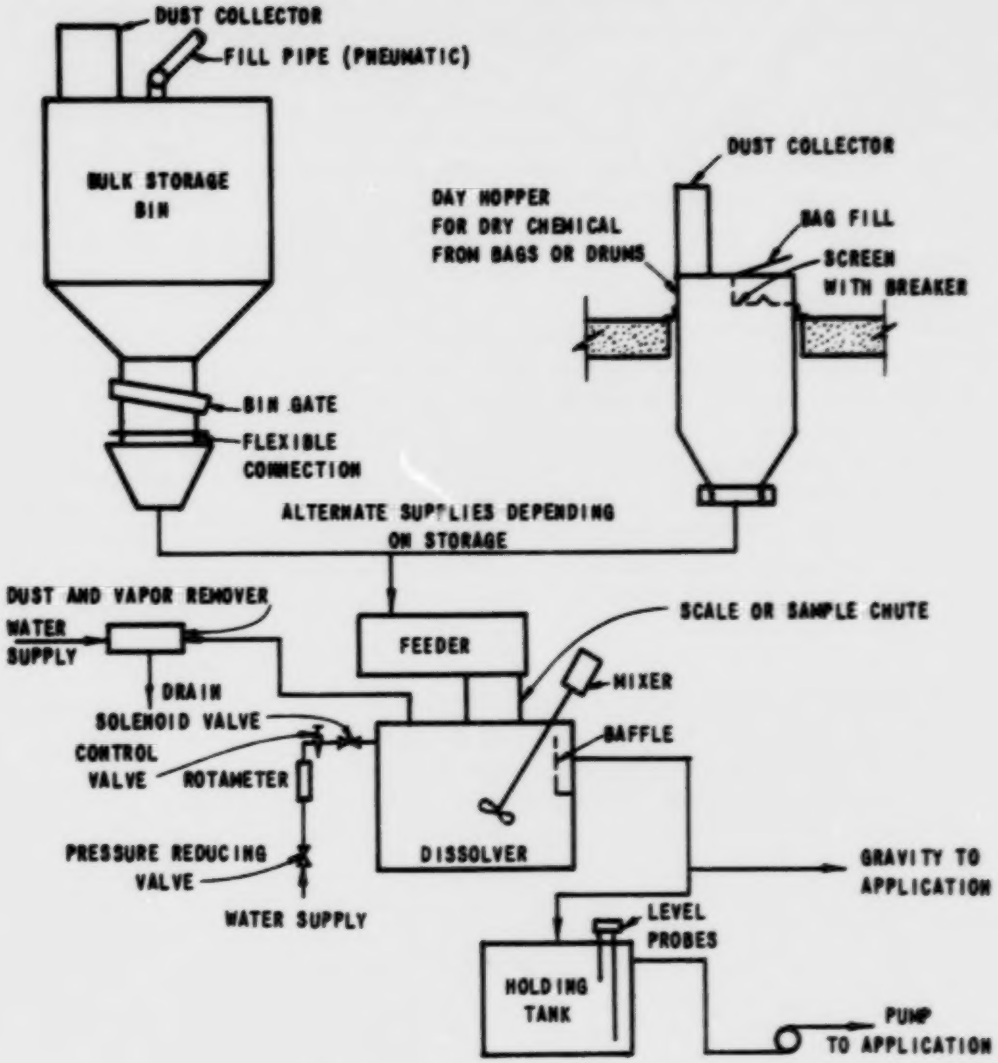

DUST COLLECTOR
FILL PIPE (PNEUMATIC)
BULK STORAGE BIN
BIN GATE
FLEXIBLE CONNECTION
DUST COLLECTOR
DAY HOPPER FOR DRY CHEMICAL FROM BAGS OR DRUMS
BAG FILL
SCREEN WITH BREAKER
ALTERNATE SUPPLIES DEPENDING ON STORAGE
DUST AND VAPOR REMOVER
WATER SUPPLY
DRAIN
FEEDER
SCALE OR SAMPLE CHUTE
MIXER
SOLENOID VALVE
CONTROL VALVE
ROTAMETER
BAFFLE
PRESSURE REDUCING VALVE
WATER SUPPLY
DISSOLVER
GRAVITY TO APPLICATION
LEVEL PROBES
HOLDING TANK
PUMP
TO APPLICATION

FIGURE 5-1

TYPICAL DRY FEED SYSTEM

Volumetric dry feeders presently in general use are of the screw type. Two designs of screw feed mechanism are available. Both allow even withdrawal across the bottom of the feeder hopper to prevent hopper dead zones. One screw design is the variable pitch type with the pitch expanding unevenly to the discharge point. The second screw design is the constant pitch type expanding evenly to the discharge point. This type of screw design is the constant pitch-reciprocating type. This type has each half of the screw turned in opposite directions so that the turning and reciprocating motion alternately fills one half of the screw while the other half of the screw is discharging. The variable pitch screw has one point of discharge, while the constant pitch-reciprocating screw has two points of discharge, one at each end of the screw. The accuracy of volumetric feeders is influenced by the character of the material being fed and ranges between ±1 percent for free-flowing materials and ± 7 percent for cohesive materials. This accuracy is volumetric and should not be related to accuracy by weight (gravimetric).

Where the greatest accuracy and the most economical use of chemical is desired, the loss-in-weight type feeder should be selected. This feeder is limited to the low and intermediate feed rates up to a maximum rate of approximately 4,000 lb hr. The loss-in-weight type feeder consists of a material hopper and feeding mechanism mounted on enclosed scales. The feed rate controller retracts the scale poise weight to deliver the dry chemical at the desired rate. The feeding mechanism must feed at this rate to maintain the balance of the scale. Any unbalance of the scale beam causes a corrective change in the output of the feeding mechanism. Continuous comparison of actual hopper weight with set hopper weight prevents cumulative errors. Accuracy of the loss-in-weight feeder is ± 1 percent by weight of the set rate.

Belt type gravimetric feeders span the capacity ranges of volumetric and loss-in-weight feeders and can usually be sized for all applications encountered in wastewater treatment applications. Initial expense is greater than for the volumetric feeder and slightly less than for the loss-in-weight feeder. Belt type gravimetric feeders consist of a basic belt feeder incorporating a weighing and control system. Feed rates can be varied by changing either the weight per foot of belt, or the belt speed, or both. Controllers in general use are mechanical, pneumatic, electric, and mechanical-vibrating. Accuracy specified for belt type gravimetric feeders should be within ± 1 percent of set rate. Materials of construction of feed equipment normally include mild steel hoppers, stainless steel mechanism components, and rubber surfaced feed belts.

Because alum solution is corrosive, dissolving or solution chambers should be constructed of type 316 stainless steel, fiberglass reinforced plastic (FRP), or plastics. Dissolvers should be sized for preparation of the desired solution strength. The solution strength usually recommended is 0.5 lb of alum to 1 gal. of water, or a 6 percent solution. The dissolving chamber is designed for a minimum detention time of 5 minutes at the maximum feed rate. Because excessive dilution may be detrimental to coagulation, eductors, or float valves that would ordinarily be used ahead of centrifugal pumps, are not recommended. Dissolvers should be equipped with water meters and mechanical mixers so that the water to alum ratio may be properly established and controlled.

5.2.1.5 Piping and Accessories

FRP, plastics (polyvinyl chloride, polyethylene, polypropylene, and other similar materials), and rubber are general use and are recommended for alum solutions. Care must be taken to provide adequate support for these piping systems, with close attention given to spans between supports so that objectionable deflection will not be experienced. The alum solution should be injected into a zone of rapid mixing or turbulent flow.

Solution flow by gravity to the point of discharge is desirable. When gravity flow is not possible, transfer components should be selected that require little or no dilution. When metering pumps or proportioning weir tanks are used, return of excess flow to a holding tank should be considered. Metering pumps are discussed further in the section on liquid alum.

Valves used in solution lines should be plastic, type 316 stainless steel or rubber-lined iron or steel.

5.2.1.6 Pacing and Control

Standard instrument control and pacing signals are generally acceptable for common feeder system operation. Volumetric and gravimetric feeders are usually adaptable to operation from any standard instrument signals.

When solution must be pumped, consideration should be given to use of holding tanks between the dry feed system and feed pumps, and the solution water supply should be controlled to prevent excessive dilution. The dry feeders may be started and stopped by tank level probes. Variable control metering pumps can then transfer the alum stock solution to the point of application without further dilution.

Means should be provided for calibration of the chemical feeders. Volumetric feeders may be mounted on platform scales. Belt feeders should include a sample shute and box to catch samples for checking actual delivery with set delivery.

Gravimetric feeders are usually furnished with totalizers only. Remote instrumentation is frequently used with gravimetric equipment, but seldom used with volumetric equipment.

5.2.2. Liquid Alum

5.2.2.1 Properties and Availability

Liquid alum is shipped in rubber-lined or stainless steel, insulated tank cars or trucks. Alum shipped during the winter is heated prior to shipment so that crystallization will not occur during transit. Liquid Alum is shipped at a solution strength of about 8.3 percent as Al_2O_3 or about 49 percent as $Al_2(SO_4)_3 \cdot 14H_2O$. The latter solution weighs about 11 lb gal at 60 °F and contains about 5.4 lb dry alum (17 percent Al_2O_3) per gal of liquid. This solution will begin to crystallize at 30 °F and freezes at about 18 °F.

Crystallization temperatures of various solution strengths are shown in Figure 5-2.

The viscosity of various alum solutions is given in Figure 5-3.

Tank truck lots of 3,000 to 5,000 gal are available. Tank car lots are available in quantities of 7,000 to 18,000 gal. Production locations of liquid alum are listed in Table 5-1. The current price range of liquid alum on an equivalent dry alum (17 percent Al_2O_3) basis is about $45 to $50 ton, F.O.B. the point of manufacture. Liquid alum will generally be more economical than dry alum if the point of use is within a 50 to 100 mile radius of the manufacturing plant.

5.2.2.2 General Design Considerations

Bulk unloading facilities usually must be provided at the treatment plant. Rail cars are constructed for top unloading and therefore require an air supply system and flexible connectors to pneumatically displace the alum from the car. U.S. Department of Transportation regulations concerning chemical tank car unloading should be observed. Tank truck unloading is usually accomplished by gravity or by a truck mounted pump.

Established practice in the treatment field has been to dilute liquid alum prior to application. However, recent studies have shown that feeding undiluted liquid alum results in better coagulation and settling. This is reportedly due to prevention of hydrolysis of the alum.

No particular industrial hazards are encountered in handling liquid alum. However, a face shield and gloves should be worn around leaking equipment. The eyes or skin should be flushed and washed upon contact with liquid alum. Liquid alum becomes very slick upon evaporation and therefore spillage should be avoided.

5.2.2.3 Storage

Liquid alum is stored without dilution at the shipping concentration. Storage tanks may be open if indoors but must be closed and vented if outdoors. Outdoor tanks should also be heated, if necessary, to keep the temperature above 45°F to prevent crystallization. Storage tanks should be constructed of type 316 stainless steel; FRP; steel lined with rubber, polyvinyl chloride, or lead. Liquid alum can be stored indefinitely without deterioration.

Storage tanks should be sized according to maximum feed rate, shipping time required, and quantity of shipment. Tanks should generally be sized for 1 1/2 times the quantity of shipments. A 10-day to 2-week supply should be provided to allow for unforeseen shipping delays.

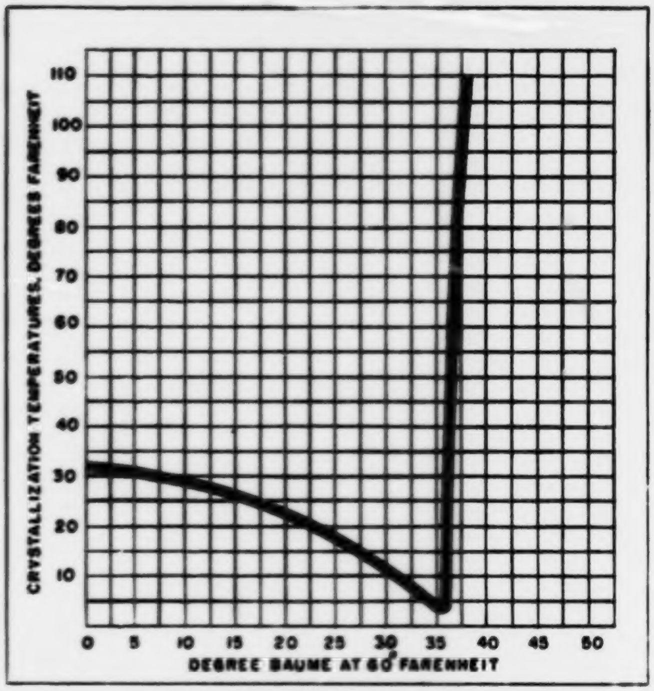


FIGURE 5-2
CRYSTALIZATION TEMPERATURES OF ALUM SOLUTIONS
(Courtesy of American Cyanamid Co.)

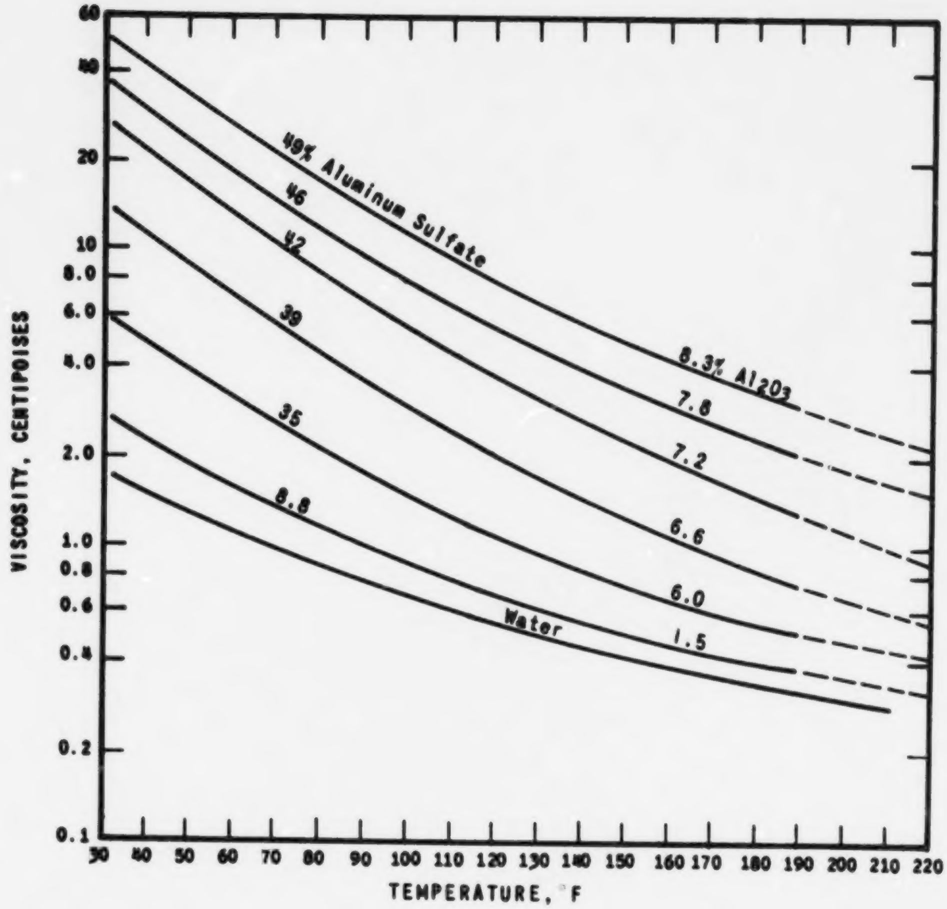


FIGURE 5-3

VISCOSITY OF ALUM SOLUTIONS
(Courtesy of Allied Chemical Co.)

5.2.2.4 Feeding Equipment

Various types of gravity or pressure feeding and metering units are available. Figures 5-4 and 5-5 illustrate commonly used feed systems. The rotodip-type feeder or rotameter is often used for gravity feed and the metering pump for pressure feed systems.

The pressure or head available at the point of application frequently determines the feeding system to be used. The rotodip feeder can be supplied from overhead storage by gravity with the use of an internal level control valve, as shown by Figure 5-4. It may also be supplied by a centrifugal pump. The latter arrangement requires an excess flow return line to the storage tank, as shown by Figure 5-5. Centrifugal pumps should be direct-connected but not close-coupled because of possible leakage into the motor, and should be constructed of type 316 stainless steel, FRP, and plastics.

Metering pumps, currently available, allow a wide range of capacity compared with the rotodip and rotameter systems. Hydraulic diaphragm type pumps are preferable to other type pumps and should be protected with an internal or external relief valve. A back pressure valve is usually required in the pump discharge to provide efficient check valve action. Materials of construction for feeding equipment should be as recommended by the manufacturer for the service, but depending on the type of system, will generally include type 316 stainless steel, FRP, plastics, and rubber.

5.2.2.5 Piping and Accessories

Piping systems for alum should be FRP, plastics (subject to temperature limits), type 316 stainless steel, or lead. Piping and valves used for alum solutions are also discussed in the preceding section on dry alum.

5.2.2.6 Pacing and Control

The feeding systems described above are volumetric, and the feeders generally available can be adapted to receive standard instrument pacing signals. The signals can be used to vary motor speed, variable-speed transmission setting, stroke speed and stroke length where applicable. A totalizer is usually furnished with a rotodip-type feeder, and remote instruments are available. Instrumentation is rarely used with rotameters and metering pumps.

5.2.3 Reactions of Aluminum Sulfate

Reactions between alum and the normal constituents of wastewaters are influenced by many factors, hence it is impossible to predict accurately the amount of alum that will react with a given amount of alkalinity, lime or soda ash which may have been added to the wastewater. Theoretical reactions can be written which will serve as a general guide, but, in general, the optimum dosage in each case must be determined by laboratory jar tests.

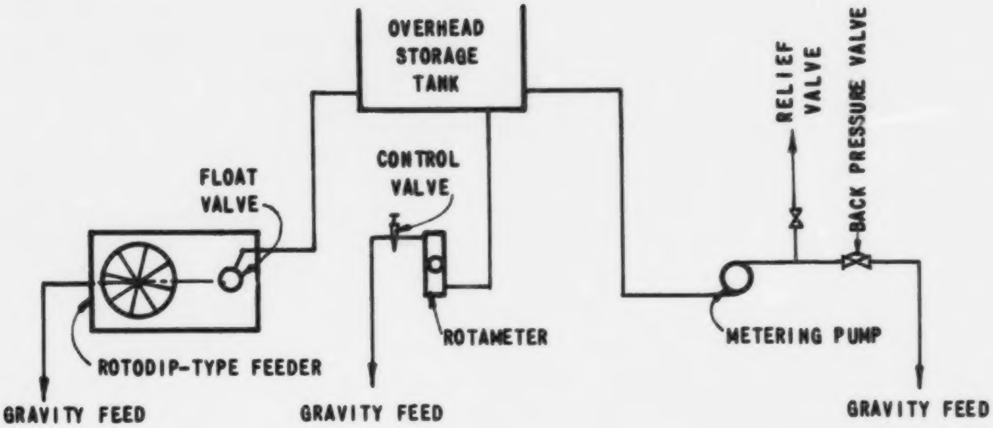

OVERHEAD
STORAGE
TANK
RELIEF
VALVE
BACK PRESSURE VALVE
CONTROL
VALVE
FLOAT
VALVE
ROTAMETER
METERING PUMP
ROTODIP-TYPE FEEDER
GRAVITY FEED
GRAVITY FEED
GRAVITY FEED

FIGURE 5-4

ALTERNATIVE LIQUID FEED SYSTEMS
FOR OVERHEAD STORAGE

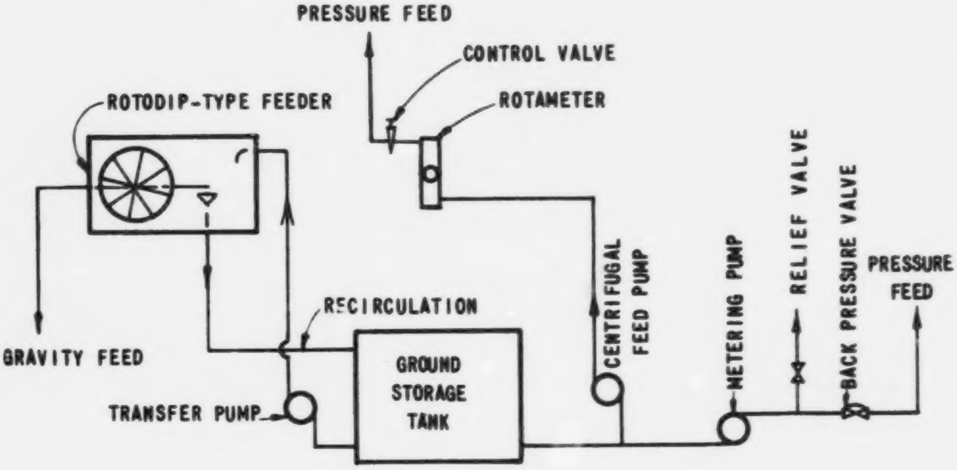

PRESSURE FEED
CONTROL VALVE
ROTODIP-TYPE FEEDER
ROTAMETER
RELIEF VALVE
BACK PRESSURE VALVE
PRESSURE
FEED
CENTRIFUGAL
FEED PUMP
METERING PUMP
RECIRCULATION
GRAVITY FEED
GROUND
STORAGE
TANK
TRANSFER PUMP

FIGURE 5-5

ALTERNATIVE LIQUID FEED SYSTEMS
FOR GROUND STORAGE

The simplest case is the reaction of Al^{3+} with OH^- ions made available by the ionization of water or by the alkalinity of the water.

Solution of alum in water produces:

$$Al_2(SO_4)_3 \rightleftharpoons 2Al^{3+} + 3\,(SO_4)^{2-}$$

Hydroxyl ions become available from ionization of water:

$$H_2O \rightleftharpoons H^{+-} + OH^-$$

The aluminum ions (Al^{3+}) then react:

$$2\,Al^{3+} + 6\,OH^- \rightleftharpoons 2\,Al\,(OH)_3$$

Consumption of hydroxyl ions will result in a decrease in the alkalinity. Where the alkalinity of the wastewater is inadequate for the alum dosage, the pH must be increased by the addition of hydrated lime, soda ash or caustic soda. The reactions of alum with the common alkaline reagents are shown in Table 5-2. While these reactions are an oversimplification of what actually takes place, they do serve to indicate orders of magnitude and some by-products of alum treatment.

TABLE 5-2

REACTIONS OF ALUMINUM SULFATE

$$Al_2\,(SO_4)_3 + 3\,Ca\,(HCO_3)_2 \longrightarrow 2\,Al\,(OH)_3\downarrow + 3\,CaSO_4 + 6\,CO_2\uparrow$$

$$Al_2\,(SO_4)_3 + 3\,Na_2CO_3 + 3\,H_2O \longrightarrow 2\,AL\,(OH)_3\downarrow + 3\,CO_2\uparrow$$

$$Al_2\,(SO_4)_3 + 3\,Ca\,(OH)_2 \longrightarrow 2\,Al\,(OH)_3\downarrow + 3\,CaSO_4$$

In terms of quantities, the reactions in Table 5-2 can be expressed as follows:

1 mg l of alum reacts with:

0.50 mg l alkalinity, expressed as $CaCO_3$

0.39 mg l 95 percent hydrated lime as $Ca\,(OH)_2$

0.54 mg l soda ash as Na_2CO_3

These approximate amounts of alkali when added to wastewater will maintain the alkalinity of the water unchanged when 1 mg l of alum is added. For example, if no alkalinity is added, 1 mg l of alum will reduce the alkalinity of 0.50 mg l as $CaCO_3$ but alkalinity can be maintained unchanged if 0.39 mg l of hydrated lime is added. This lowering of natural alkalinity is desirable in many cases to attain the pH range for optimum coagulation.

For each mg l of alum dosage, the sulfate (SO_4) content of the water will be increased approximately 0.49 mg l and the CO_2 content of the water will be increased approximately 0.44 mg l.

5.3 Iron Compounds

Iron compounds have pH coagulation ranges and floc characteristics similar to aluminum sulfate. The cost of iron compounds may often be less than the cost of alum. However, the iron compounds are generally corrosive and often present difficulties in dissolving, and their use may result in high soluble iron concentrations in process effluents.

5.3.1 Liquid Ferric Chloride

5.3.1.1 Properties and Availability

Liquid ferric chloride is a corrosive, dark brown oily-appearing solution having a weight as shipped and stored of 11.2 to 12.4 lb gal (35 to 45 percent $FeCl_3$) (1). The ferric chloride content of these solutions, as $FeCl_3$, is 3.95 to 5.58 lb gal. Shipping concentrations vary from summer to winter due to the relatively high crystallization temperature of the more concentrated solutions as shown by Figure 5-6. The pH of a 1 percent solution is 2.0.

The molecular weight of ferric chloride is 162.22. Viscosities of ferric chloride solutions at various temperatures are presented in Figure 5-7.

Liquid ferric chloride is shipped in 3,000 to 4,000 gal bulk truckload lots, in 4,000 to 10,000 gal bulk carload lots, and in 5 and 13 gal carboys. Liquid ferric chloride is produced at the following locations (2):

Dow Chemical Co.
Midland, Michigan

Pennwalt Corp.
Philadelphia, Pa. (Plant at Wyandotte, Mich.)

The current price of liquid ferric chloride in bulk quantities is about $0.04 to $0.05 lb (as $FeCl_3$), F.O.B. the point of manufacture.

Tank trucks and cars are normally unloaded pneumatically, and operating procedures must be closely followed to avoid spills and accidents. The safety vent cap and assembly (painted red) should be removed prior to opening the unloading connection to depressurize the tank car or truck, prior to unloading.

5.3.1.2 General Design Considerations

Ferric chloride solutions are corrosive to many common materials and cause stains which are difficult to remove. Areas which are subject to staining should be protected with resistant paint or rubber mats.

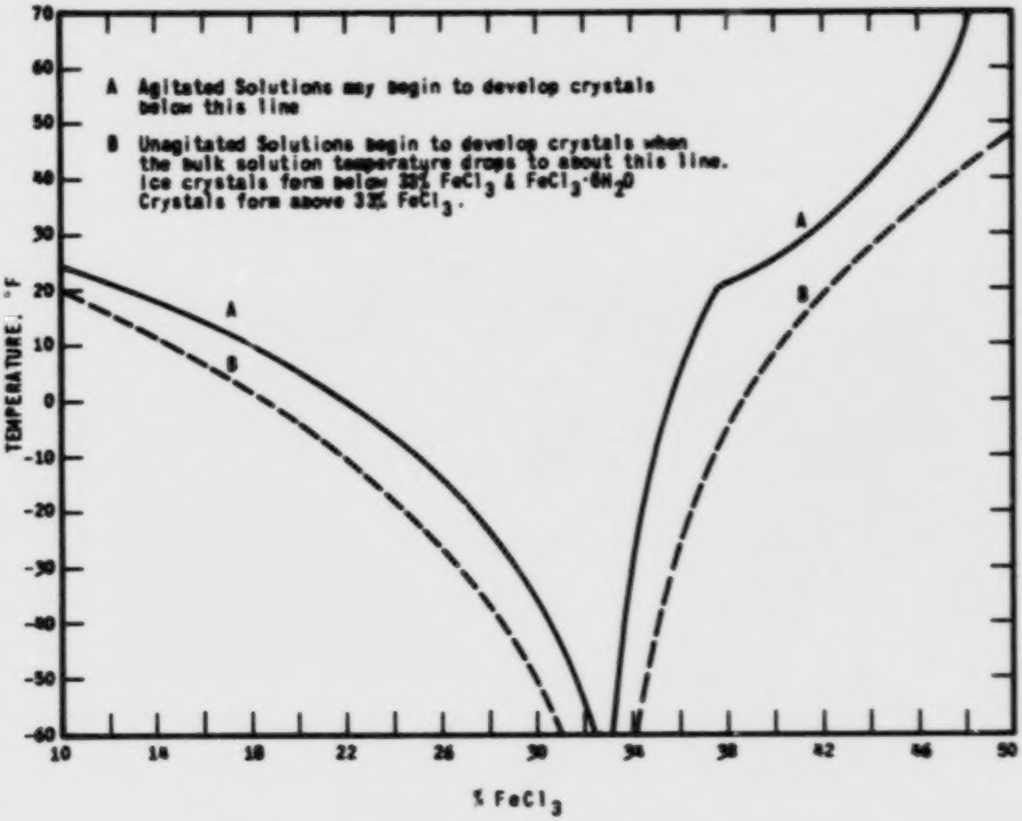


FIGURE 5-6

FREEZING POINT CURVES FOR COMMERCIAL FERRIC CHLORIDE SOLUTIONS
(Courtesy of Dow Chemical Co.)

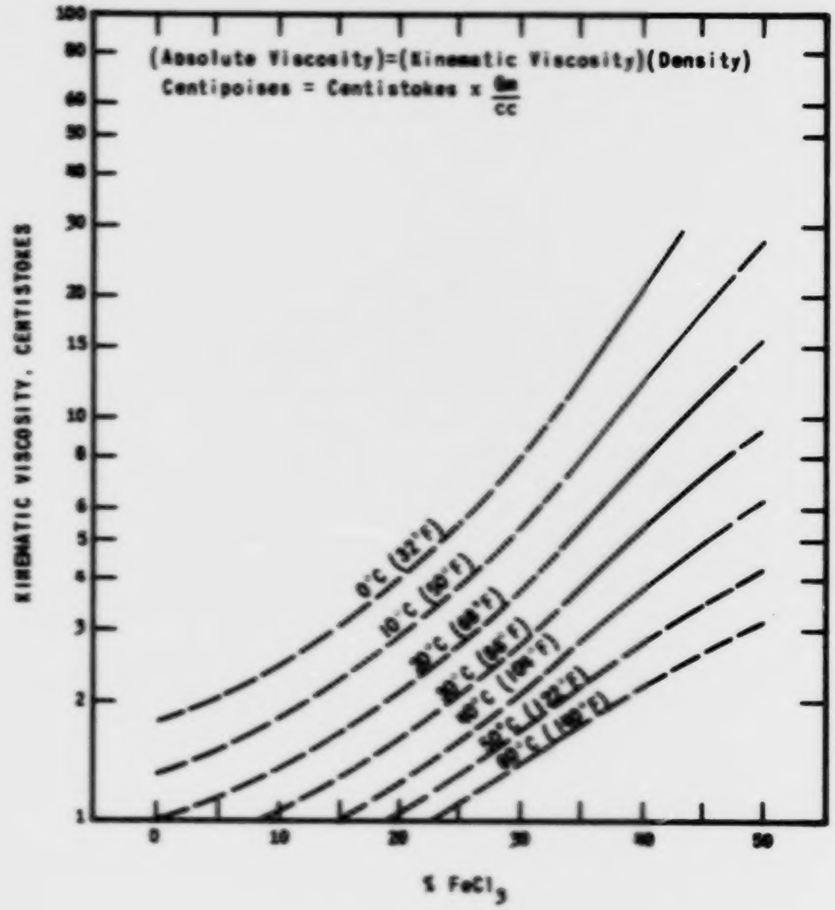


FIGURE 5-7

VISCOSITY VS COMPOSITION OF FERRIC CHLORIDE SOLUTIONS AT VARIOUS TEMPERATURES
(Courtesy of Dow Chemical Co.)

Normal precautions should be employed when cleaning ferric chloride handling equipment. Workmen should wear rubber gloves, rubber apron, and goggles or a face shield. If ferric chloride comes in contact with the eyes or skin, flush with copious quantities of running water and call a physician. If ferric chloride is ingested, induce vomiting and call a physician.

5.3.1.3 Storage

Ferric chloride solution can be stored as shipped. Storage tanks should have a free vent or vacuum relief valve. Tanks may be constructed of FRP, rubber lined steel, of plastic lined steel. Resin-impregnated carbon or graphite are also suitable materials for storage containers.

It may be necessary in most instances to house liquid ferric chloride tanks in heated areas or provide tank heaters or insulation to prevent crystallization. Ferric chloride can be stored for long periods of time without deterioration. The total storage capacity should be $1\frac{1}{2}$ times the largest anticipated shipment, and should provide at least a 10-day to 2-week supply of the chemical at the design average dosage.

5.3.1.4 Feeding Equipment

Feeding equipment and systems described for liquid alum generally apply to ferric chloride except for materials of construction and the use of glass tube rotameters.

It may not be desirable to dilute the ferric chloride solution from its shipping concentration to a weaker feed solution because of possible hydrolysis. Ferric chloride solutions may be transferred from underground storage to day tanks with impervious graphite or rubber lined self-priming centrifugal pumps having teflon rotary and stationary seals. Because of the tendency for liquid ferric chloride to stain or deposit, glass-tube rotameters should not be used for metering this solution. Rotodip feeders and diaphragm metering pumps are often used for ferric chloride, and should be constructed of materials such as rubber-lined steel and plastics.

5.3.1.5 Piping and Accessories

Materials for piping and transporting ferric chloride should be rubber or Saran-lined steel, hard rubber, FRP, or plastics. Valving should consist of rubber or resin-lined diaphragm valves, Saran-lined valves with teflon diaphragms, rubber-sleeved pinch-type valves, or plastic ball valves. Gasket material for large openings such as manholes in storage tanks should be soft rubber; all other gaskets should be graphite-impregnated blue asbestos, teflon, or vinyl.

5.3.1.6 Pacing and Control

System pacing and control requirements are similar to those discussed previously for liquid alum.

5.3.2 Ferrous Chloride (Waste Pickle Liquor)

5.3.2.1 Properties and Availability

Ferrous chloride, $FeCl_2$, as a liquid is available in the form of waste pickle liquor from steel processing. The liquor weighs between 9.9 and 10.4 lb/gal and contains 20 to 25 percent $FeCl_2$ or about 10 percent available Fe^{2+}. A 22 percent solution of $FeCl_2$ will crystallize at a temperature of -4 °F. The molecular weight of $FeCl_2$ is 126.76. Free acid in waste pickle liquor can vary from 1 to 10 percent and usually averages about 1.5 to 2.0 percent. Ferrous chloride is slightly less corrosive than ferric chloride.

Waste pickle liquor is available in 4,000 gal truckload lots and a variety of carload lots. In most instances the availability of waste pickle liquor will depend on the proximity to steel processing plants. Dow Chemical Company produces a waste pickle liquor, having an $FeCl_2$ content of about 22 percent at a price of $0.04/lb of $FeCl_2$ in bulk car or truckload quantities, F.O.B. Midland, Michigan.

5.3.2.2 General Design Considerations

Since ferrous chloride or waste pickle liquor may not be available on a continuous basis, storage and feeding equipment should be suitable for handling ferric chloride. Therefore, the ferric chloride section should be referred to for storage and handling details.

5.3.3 Ferric Sulfate

5.3.3.1 Properties and Availability

Ferric sulfate is marketed as dry, partially-hydrated granules with the formula $Fe_2(SO_4)_3 \cdot X\ H_2O$, where X is approximately 7. Typical properties of one commercial product (2) are presented below:

Molecular Weight	526
Bulk Density	56-60 lb/cu ft
Water Soluble Iron Expressed as Fe	21.5 percent
Water Soluble Fe^{+3}	19.5 percent
Water Soluble Fe^{+2}	2.0 percent
Insolubles Total	4.0 percent
Free Acid	2.5 percent
Moisture @ 105°C.	2.0 percent

Ferric sulfate is shipped in car and truck load lots of 50 lb and 100 lb moistureproof paper bags and 200 lb and 400 lb fiber drums. Bulk carload shipments in box and closed hopper cars are available. The major producer is Cities Service Company, with a plant located at Copper Hill, Tennessee.

The current price of ferric sulfate (21.8 percent Fe) is about $39 ton, F.O.B. Copper Hill, Tennessee. The bagged form costs from $6 to $11 ton more than the bulk.

General precautions should be observed when handling ferric sulfate, such as wearing goggles and dust masks, and areas of the body that come in contact with the dust or vapor should be washed promptly.

5.3.3.2 General Design Considerations

Aeration of ferric sulfate should be held to a minimum because of the hygroscopic nature of the material, particularly in damp atmospheres. Mixing of ferric sulfate and quicklime in conveying and dust vent systems should be avoided as caking and excessive heating can result. The presence of ferric sulfate and lime in combination has been known to destroy cloth bags in pneumatic unloading devices (3). Because ferric sulfate in the presence of moisture will stain, precautions similar to those discussed for ferric chloride should be observed.

5.3.3.3 Storage

Ferric sulfate is usually stored in the dry state either in the shipping bags or in bulk in concrete or steel bins. Bulk storage bins should be as tight as possible to avoid moisture absorption, but dust collector vents are permissible and desirable. Hoppers on bulk storage bins should have a minimum slope of 36° however, a greater angle is prefered.

Bins may be located inside or outside and the material transferred by bucket elevator, screw or air conveyors. Ferric sulfate stored in bins usually absorbs some moisture and forms a thin protective crust which retards further absorption until the crust is broken.

5.3.3.4 Feeding Equipment

Feed solutions are usually made up at a water to chemical ratio of 2:1 to 8:1 (on a weight basis) with the usual ratio being 4:1 with a 20-minute detention time. Care must be taken not to dilute ferric sulfate solutions to less than 1 percent to prevent hydrolysis and deposition of ferric hydroxide. Ferric sulfate is actively corrosive in solution, and dissolving and transporting equipment should be fabricated of type 316 stainless steel, rubber, plastics, ceramics or lead.

Dry feeding requirements are similar to those for dry alum except that belt type feeders are rarely used because of their open type of construction. Closed construction, as found in the volumetric and loss-in-weight-type feeders, generally exposes a minimum of operating components to the vapor, and thereby minimizes maintenance. A water jet vapor remover should be provided at the dissolver to protect both the machinery and operator.

5.3.3.5 Piping and Accessories

Piping systems for ferric sulfate should be FRP, plastics, type 316 stainless steel, rubber or glass.

5.3.3.6 Pacing and Control

System pacing and control are the same as discussed for dry alum.

5.3.4 Ferrous Sulfate

5.3.4.1 Properties and Availability

Ferrous sulfate or copperas is a byproduct of pickling steel and is produced as granules, crystals, powder, and lumps. The most common commercial form of ferrous sulfate is $FeSO_4 \cdot 7H_2O$, with a molecular weight of 278, and containing 55 to 58 percent $FeSO_4$ and 20 to 21 percent Fe. The product has a bulk density of 62 to 66 lb/cu ft. When dissolved, ferrous sulfate is acidic. The composition of ferrous sulfate may be quite variable and should be established by consulting the nearest manufacturers.

Bulk, drum (400 lb) and bag (50 and 100 lb) shipments are available from producers at the following locations:

American Cyanamid Co.	Savannah, Georgia
Byproducts Processing Co., Inc.	Baltimore, Maryland
Glidden Co.	Baltimore, Maryland
Cosmin Corp.	Baltimore, Maryland
NL Industries	St. Louis, Missouri
NL Industries	Sayreville, New Jersey

The current price of ferrous sulfate in bulk carload and truckload quantities is about $18/ton (21 percent Fe). The bagged cost is $24/ton.

Ferrous sulfate is also available in a wet state in bulk form from some plants. This form is likely to be difficult to handle and the manufacturer should be consulted for specific information and instructions.

Dry ferrous sulfate cakes at storage temperatures above 68°F, is efflorescent in dry air, and oxidizes and hydrates further in moist air.

General precautions similar to those for ferric sulfate, with respect to dust and handling acidic solutions, should be observed when working with ferrous sulfate. Mixing quicklime and ferrous sulfate produces high temperatures and the possibility of fire.

5.3.4.2 General Design Considerations

The granular form of ferrous sulfate has the best feeding characteristics and gravimetric or volumetric feeding equipment may be used.

The optimum chemical to water ratio for continuous dissolving is 0.5 lb/gal. of 6 percent with a detention time of 5 minutes in the dissolver. Mechanical agitation should be provided in the dissolver to assure complete solution. Lead, rubber, iron, plastics, and type 304 stainless steel can be used as construction materials for handling solutions of ferrous sulfate.

Storage, feeding and transporting systems probably should be suitable for handling ferric sulfate as an alternative to ferrous sulfate.

5.3.5 Reactions of Iron Compounds

Ferric sulfate and ferric chloride react with the alkalinity of wastewater or with the added alkaline materials such as lime or soda ash. The reactions may be written to show precipitation of ferric hydroxide, although in practice, as with alum, the reactions are more complicated than this. The reactions are shown in Table 5-3, using ferric sulfate.

TABLE 5-3

REACTIONS OF FERRIC SULFATE

$$Fe_2(SO_4)_3 + 3\,Ca(HCO_3)_2 \longrightarrow 2\,Fe(OH)_3\downarrow + 3\,CaSO_4 + 6\,CO_2\uparrow$$

$$Fe_2(SO_4)_3 + 3\,Na_2CO_3 + 3\,H_2O \longrightarrow 2\,Fe(OH)_3\downarrow + 3\,Na_2SO_4 + 3\,CO_2\uparrow$$

$$Fe_2(SO_4)_3 + 3\,Ca(OH)_2 \longrightarrow 2\,Fe(OH)_3\downarrow + 3\,CaSO_4$$

Ferric chloride can be substituted in these reactions.

In terms of useful quantities, the reactions of Table 5-3 can be expressed as follows:

1. 1 mg/l of $Fe_2(SO_4)_3 \cdot 7H_2O$ reacts with:

   0.57 mg/l alkalinity, expressed as $CaCO_3$

   0.44 mg/l 95 percent hydrated lime as $Ca(OH)_2$

   0.62 mg/l soda ash as Na_2CO_3

2. 1 mg/l of anhydrous $FeCl_3$ reacts with:

   0.92 mg/l alkalinity expressed as $CaCO_3$

   0.72 mg/l 95 percent hydrated lime as $Ca(OH)_2$

   1.00 mg/l soda ash as Na_2CO_3

Ferrous sulfate and ferrous chloride react with the alkalinity of wastewater or with the added alkaline materials such as lime to precipitate ferrous hydroxide. The ferrous hydroxide is oxidized to ferric hydroxide by dissolved oxygen in wastewater. Typical reactions are shown in Table 5-4, using ferrous sulfate.

TABLE 5-4

REACTIONS OF FERROUS SULFATE

$$FeSO_4 + Ca(HCO_3)_2 \longrightarrow Fe(OH)_2\downarrow + Ca\,SO_4 + 2CO_2\uparrow$$

$$FeSO_4 + Ca(OH)_2 \longrightarrow Fe(OH)_2\downarrow + Ca\,SO_4$$

$$4\,Fe(OH)_2 + O_2 + 2H_2 \longrightarrow 4\,Fe\,(OH)_3\downarrow$$

Ferrous hydroxide is rather soluble and oxidation to the more insoluble ferric hydroxide is necessary if high iron residuals in effluents are to be avoided. Flocculation with ferrous iron is improved by addition of lime or caustic soda at a rate of 1 to 2 mg mg Fe to serve as a floc conditioning agent. Polymers are also generally required to produce a clear effluent.

5.4 Lime

The term "lime" applies to a variety of chemicals which are alkaline in nature and contain principally calcium, oxygen and, in some cases, magnesium. In this grouping are included quicklime, dolomitic lime, hydrated lime, dolomitic hydrated lime, limestone, and dolomite. This section is restricted to discussion of quicklime and hydrated lime, but the dolomitic counterparts of these chemicals, i.e., the high-magnesium forms, are quite applicable for wastewater treatment and are generally similar in physical requirements.

5.4.1 Quicklime

5.4.1.1 Properties and Availability

Quicklime, CaO, has a density range of approximately 55 to 75 lb cu ft, and a molecular weight of 56.08. A slurry for feeding, called milk of lime, can be prepared with up to 45 percent solids. Lime is only slightly soluble, and both lime dust and slurries are caustic in nature. A saturated solution of lime has a pH of about 12.4.

Lime can be purchased in bulk in both car and truck load lots. It is also shipped in 80 and 100 lb multiwall "moistureproof" paper bags. Lime is produced at the locations indicated by Table 5-5.

TABLE 5-5

PARTIAL LIST OF LIME MANUFACTURING PLANTS (4)

Location	Manufacturer	Form of Lime Available
ALABAMA		
Allgood	Cheney Lime & Cement Co.	High Calcium
Keystone	Southern Cement Co. Div. Martin Marietta Corp.	High Calcium
Landmark	Cheney Lime & Cement Co.	High Calcium
Montevallo	U.S. Gypsum Co.	High Calcium
Roberta	Southern Cement Co. Div. Martin Marietta Corp.	High Calcium
Saginaw	Longview Lime Co., Div. Woodward Co., Div. Mead Corp.	High Calcium
Siluria	Alabaster Lime Co	High Calcium
ARIZONA		
Douglas	Paul Lime Plant, Inc.	High Calcium
Globe	Hoopes & Co.	High Calcium
Nelson	U.S. Lime Div., The Flintkote Co.	High Calcium
ARKANSAS		
Batesville	Batesville White Lime Co., Div. Rangaire Corp.	High Calcium
CALIFORNIA		
City of Industry	U.S. Lime Div., The Flintkote Co.	High Calcium
Diamond Springs	Diamond Springs Lime Co.	High Calcium
Lucerne Valley	Pfizer, Inc., Minerals, Pigments and Metals Div.	High Calcium
Richmond	U.S. Lime Div., The Flintkote Co.	High Calcium
Salinas	Kaiser Aluminum & Chemical Corp. (currently captive lime)	Dolomitic
Sonora	U.S. Lime Div., The Flintkote Co.	Dolomitic
Westend	Stauffer Chemical Co.	High Calcium
COLORADO		
Ft. Morgan	Great Western Sugar Co.	High Calcium
CONNECTICUT		
Canaan	Pfizer, Inc., Minerals, Pigments and Metals Div.	Dolomitic

TABLE 5-5 (continued)

Location	Manufacturer	Form of Lime Available
FLORIDA		
Brooksville	Chemical Lime Co.	High Calcium
Sumterville	Dixie Lime and Stone Co.	High Calcium
ILLINOIS		
Marblehead	Marblehead Lime Co.	High Calcium
McCook	Standard Lime & Refractories Div., Martin Marietta Corp.	Dolomitic
Quincy	Marblehead Lime Co.	High Calcium
So. Chicago	Marblehead Lime Co.	High Calcium
Thornton	Marblehead Lime Co.	Dolomitic
INDIANA		
Buffington	Marblehead Lime Co.	High Calcium
IOWA		
Davenport	Linwood Stone Products Co., Inc.	High Calcium
KENTUCKY		
Carntown	Black River Mining Co.	High Calcium
LOUISIANA		
Morgan City	Pelican State Lime Corp.	High Calcium
New Orleans	U.S. Gypsum Co.	High Calcium
MARYLAND		
LeGore	LeGore Lime Co.	High Calcium
Woodsboro	S.W. Barrick & Sons, Inc.	High Calcium
MASSACHUSETTS		
Adams	Pfizer, Inc., Minerals, Pigments and Metals Div.	High Calcium
Lee	Lee Lime Corp.	Dolomitic
MICHIGAN		
Detroit	Detroit Lime Co.	High Calcium
Ludington	Dow Chemical Co. (currently captive lime)	High Calcium
Menominee	Limestone Products Co., Div. C. Reiss Coal Co.	High Calcium
River Rouge	Marblehead Lime Co.	High Calcium
MINNESOTA		
Duluth	Cutler Magner Co.	High Calcium

TABLE 5-5 (continued)

Location	Manufacturer	Form of Lime Available
MISSOURI		
Bonne Terre	Valley Dolomite Co.	Dolomitic
Hannibal	Marblehead Lime Co.	High Calcium
Ste. Genevieve	Mississippi Lime Co.	High Calcium
Springfield	Ash Grove Cement Co.	High Calcium
NEVADA		
Apex	U.S. Lime Div., The Flintkote Co.	High Calcium
Henderson	U.S. Lime Div., The Flintkote Co.	Dolomitic & High Calcium
McGill	Morrison-Weatherly Corp.	High Calcium
Sloan	U.S. Lime Div., The Flintkote Co.	Dolomitic & High Calcium
NEW JERSEY		
Newton	Limestone Products Corp. of America	High Calcium
OHIO		
Ashtabula	Union Carbide Olefins Co.	High Calcium
Carey	National Lime & Stone Co.	Dolomitic
Cleveland	Cuyahoga Lime Co.	High Calcium
Delaware	Marble Cliff Quarries Co.	High Calcium
Geona	U.S. Gypsum Co.	Dolomitic
Gibsonburg (2 plants)	Pfizer, Inc., Minerals, Pigments and Metal Div., National Gypsum Co.	Dolomitic
Huron	Huron Lime Co.	High Calcium
Marble Cliff	Marble Cliff Quarries Co.	High Calcium
Millersville	J. E. Baker Co.	Dolomitic
Woodville	Ohio Lime Co., Standard Lime & Refractories Div., Martin Marietta Corp.	Dolomitic
OKLAHOMA		
Marble City		
Sallisaw	St. Clair Lime Co.	High Calcium
	St. Clair Lime Co.	High Calcium
OREGON		
Baker		
Portland	Chemical Lime Co. of Oregon	High Calcium
	Ash Grove Cement Co.	High Calcium

TABLE 5-5 (continued)

Location	Manufacturer	Form of Lime Available
PENNSYLVANIA		
Annville	Bethlehem Mines Corp.	High Calcium
Bellefonte (2 plants)	National Gypsum Co., Warner Co.	High Calcium
Branchton	Mercer Lime & Stone Co.	High Calcium
Devault	Warner Co.	Dolomitic
Everett	New Enterprise Stone & Lime Co.	High Calcium
Pleasant Gap	Standard Lime & Refractories Div., Martin Marietta Corp.	High Calcium
Plymouth Meeting	G. & W. H. Corson, Inc.	Dolomitic
SOUTH DAKOTA		
Rapid City	Pete Lien & Sons, Inc	High Calcium
TENNESSEE		
Knoxville (2 plants)	Foote Mineral Co., Williams Lime Manufacturing Co.	High Calcium
TEXAS		
Blum	Round Rock Lime Companies	High Calcium
Cleburne	Texas Lime Co., Div. Rangaire Corp.	High Calcium
Clifton	Clifstone Lime Co.	High Calcium
Houston	U. S. Gypsum Co.	High Calcium
McNeil	Austin White Lime Co.	High Calcium
New Braunfels	U. S. Gypsum Co.	High Calcium
Round Rock	Round Rock Lime Companies	High Calcium
San Antonio	McDonough Bros., Inc.	High Calcium
UTAH		
Grantsville	U. S. Lime Div., The Flintkote Co.	Dolomitic & High Calcium
Lehi	Rollins Mining Supplies Co.	High Calcium
VERMONT		
Winooski	Vermont Assoc. Lime Industries, Inc.	High Calcium
VIRGINIA		
Clearbrook	W.S. Frey Co., Inc.	High Calcium
Kimballton (2 plants)	Foote Mineral Co., National Gypsum Company	High Calcium

TABLE 5-5 (continued)

Location	Manufacturer	Form of Lime Available
Stephens City	M.J. Grove Lime Co., Div. The Flintkote Co.	High Calcium
Strasburg	Chemstone Corp.	High Calcium
WASHINGTON		
Tacoma	Domtar Chemicals Inc.	High Calcium
WEST VIRGINIA		
Millville	Standard Lime & Refractories Div., Martin Marietta Corp.	Dolomitic
Riverton	Germany Valley Limestone Div., Greer Steel Co.	High Calcium
WISCONSIN		
Eden	Western Lime & Cement Co.	Dolomitic
Green Bay	Western Lime & Cement Co.	High Calcium
Knowles	Western Lime & Cement Co.	Dolomitic
Manitowoc	Rockwell Lime Co.	Dolomitic
Superior	Cutler-LaLiberte-McDougall Corp.	High Calcium

Current prices for bulk pebble quicklime range from $18/ton to $21/ton with the higher prices generally in the far west, and higher than average in the north. Bagging adds approximately $4/ton to the cost.

The CaO content of commercially available quicklime can vary quite widely over an approximate range of 70 to 96 percent. Content below 88 percent is generally considered below standard in the municipal use field (5). Purchase contracts are often based on 90 percent CaO content with provisions for payment of a bonus for each 1 percent over and a penalty for each 1 percent under the standard. A CaO content less than 75 percent probably should be rejected because of excessive grit and difficulties in slaking.

Workmen should wear protective clothing and goggles to protect the skin and eyes, as lime dust and hot slurry can cause severe burns. Areas contacted by lime should be washed immediately. Lime should not be mixed with chemicals which have water of hydration. The lime will be slaked by the water of hydration causing excessive temperature rise and possibly explosive conditions. Conveyors and bins used for more than one chemical should be thoroughly cleaned before switching chemicals.

5.4.1.2 General Design Considerations

Pebble quicklime, all passing a 3/4 in. screen and not more than 5 percent passing a No. 100 screen, is normally specified because of easier handling and less dust. Hopper agitation is generally not required with the pebble form. Published slaker capacity ratings require "soft or normally burned" limes which provide fast slaking and temperature rise, but poorer grades of limes may also be satisfactorily slaked by selection of the appropriate slaker retention time and capacity.

5.4.1.3 Storage

Storage of bagged lime should be in a dry place, and preferably elevated on pallets to avoid absorption of moisture. System capacities often make the use of bagged quicklime impractical. Maximum storage period is about 60 days.

Bulk lime is stored in air-tight concrete or steel bins having a 55 to 60 deg slope on the bin outlet. Bulk lime can be conveyed by conventional bucket elevators and screw, belt, apron, drag-chain, and bulk conveyors of mild steel construction. Pneumatic conveyors subject the lime to air-slaking and particle sizes may be reduced by attrition. Dust collectors should be provided on manually and pneumatically-filled bins.

5.4.1.4 Feeding Equipment

A typical lime storage and feed system is illustrated in Figure 5-8. Quicklime feeders are usually limited to the belt or loss-in-weight gravimetric types because of the wide variation of the bulk density. Feed equipment should have an adjustable feed range of at least 20:1 to match the operating range of the associated slaker. The feeders should have an over-under feed rate alarm to immediately warn of operation beyond set limits of control. The feeder drive should be instrumented to be interrupted in the event of excessive temperature in the slaker compartment.

Lime slakers for wastewater treatment should be of the continuous type, and the major components should include one or more slaking compartments, a dilution compartment, a grit separation compartment and a continuous grit remover. Commercial designs vary in regard to the combination of water to lime, slaking temperature, and slaking time, in obtaining the "milk of lime" suspensions.

The "paste-type" slaker admits water as required to maintain a desired mixing viscosity. This viscosity therefore sets the operating retention time of the slaker. The paste slaker usually operates with a low water to lime ratio (approximately 2:1 by weight), elevated temperature, and five-minute slaking time at maximum capacity.

The "detention" type slaker admits water to maintain a desired ratio with the lime, and therefore the lime feed rate sets the retention time of the slaker. The detention slaker operates with a wide range of water to lime ratios (2.5:1 and 6:1), moderate temperature, and a 10 minute slaking time at maximum capacity. A water to lime ratio of from 3.5:1 to 4:1 is most often used. The operating temperature in lime slakers is a function of the water to lime ratio, lime quality, heat transfer, and water temperature. Lime slaking evolves heat in hydrating the CaO to $Ca(OH)_2$ and therefore, vapor removers are required for feeder protection.

5.4.1.5 Piping and Accessories

Lime slurry should be transported by gravity in open channels wherever possible. Piping channels, and accessories may be rubber, iron, steel, concrete, and plastics. Glass tubing, such as that in rotameters, will cloud rapidly and therefore should not be used. Any abrupt directional changes in piping should include plugged tees or crosses to allow rodding-out of deposits. Long sweep elbows should be provided to allow the piping to be cleaned by the use of a cleaning "pig". Daily cleaning is desirable.

Milk of lime transfer pumps should be of the open impeller centrifugal type. Pumps having an iron body and impeller with bronze trim are suitable for this purpose. Rubber-lined pumps with rubber-covered impellers are also frequently used. Make-up tanks are usually provided ahead of centrifugal pumps to ensure a flooded suction at all times. "Plating-out" of lime is minimized by the use of soft water in the make-up tank and slurry recirculation. Turbine pumps and eductors should be avoided in transferring milk of lime because of scaling problems.

5.4.1.6 Pacing and Control

Lime slaker water proportioning is integrally-controlled or paced from the feeder. Therefore, the feeder-slaker system will follow pacing controls applied to the feeder only. As discussed previously, gravimetric feeders are adaptable to receive most standard instrumentation pacing signals. Systems can be instrumented to allow remote pacing with telemetering of temperature and feed rate to a central panel for control purposes.

The lime feeding system may be controlled by an instrumentation system integrating both plant flow and pH of the wastewater after lime addition. However, it should be recognized that pH probes require daily maintenance in this application to monitor the pH accurately. Deposits tend to build up on the probe and necessitate frequent maintenance. The low pH lime treatment systems (pH 9.5 to 10.0) can be more readily adapted to this method of control than high-lime treatment systems (pH 11.0 or greater) because less maintenance of the pH equipment is required. In a closed-loop pH-flow control system, milk of lime is prepared on a batch basis and transferred to a holding tank with variable output feeders set by the flow and pH meters to proportion the feed rate. Figure 5-8 illustrates such a control system.

5.4.2 Hydrated Lime

5.4.2.1 Properties and Availability

Hydrated lime, $Ca(OH)_2$, is usually a white powder (200 to 400 mesh); has a bulk density of 20 to 50 lb/cu ft; contains 82 to 98 percent $Ca(OH)_2$; is slightly hydroscopic; tends to flood the feeder, and will arch in storage bins if packed. The molecular weight is 74.08. The dust and slurry of hydrated lime are caustic in nature. The cost of bulk hydrated lime varies from $18 to $22/ton. Bagged lime is available but increases the cost from $4 to $16/ton. The availability of hydrated lime may be determined by contacting manufacturers listed in

Table 5-5. The pH of a saturated, hydrated lime solution is the same as that given for quicklime.

5.4.2.2 General Design Considerations

Hydrated lime is slaked lime and needs only enough water added to form milk of lime. Wetting or dissolving chambers are usually designed to provide 5 minutes detention with a ratio of 0.5 lb/gal of water or 6 percent slurry at the maximum feed rate. Hydrated lime is usually used where maximum feed rates do not exceed 250 lb/hr., i.e., smaller plants. Hydrated lime and milk of lime will irritate the eyes, nose, and respiratory system and will dry the skin. Affected areas should be washed with water.

5.4.2.3 Storage

Information given for quicklime also applies to hydrated lime except that bin agitation must be provided. Bulk bin outlets should be provided with non-flooding rotary feeders. Hopper slopes vary from 60 to 66 deg.

5.4.2.4 Feed Equipment

Volumetric or gravimetric feeders may be used, but volumetric feeders are usually selected only for installations where comparatively low feed rates are required. Dilution does not appear to be important, therefore, control of the amount of water used in the feeding operation is not considered necessary. Inexpensive hydraulic jet agitation may be furnished in the wetting chamber of the feeder as an alternative to mechanical agitation. The jets should be sized for the available water supply pressure to obtain proper mixing.

5.4.2.5 Piping and Accessories

Piping and accessories as described for quicklime are also appropriate for hydrated lime.

5.4.2.6 Pacing and Controls

Controls as listed for dry alum apply to hydrated lime. Hydraulic jets should operate continuously and only shut off when the feeder is taken out of service. Control of the feed rate with pH as well as pacing with the plant flow may be used with hydrated lime as well as quicklime.

5.4.3 Reactions of Lime

Lime is somewhat different from the hydrolyzing coagulants. When added to wastewater it increases pH and reacts with the carbonate alkalinity to precipitate calcium carbonate. If sufficient lime is added to reach a high pH, approximately 10.5, magnesium hydroxide is also precipitated. This latter precipitation enhances clarification due to the flocculant nature of the $Mg(OH)_2$. Excess calcium ions at high pH levels may be precipitated by the addition of soda ash. The above reactions are shown in Table 5-6.

TABLE 5-6

REACTIONS OF LIME

$$Ca(OH)_2 + Ca(HCO_3)_2 \longrightarrow 2\,CaCO_3\downarrow + 2H_2O$$

$$2\,Ca(OH)_2 + Mg(HCO_3)_2 \longrightarrow 2\,CaCO_3\downarrow + Mg(OH)_2\downarrow + 2H_2O$$

$$Ca(OH)_2 + Na_2CO_3 \longrightarrow CaCO_3\downarrow + 2\,NaOH$$

Reduction of the resulting high pH levels may be accomplished in one or two stages. The first stage of the two-stage method results in the precipitation of calcium carbonate through the addition of carbon dioxide according to the following reaction:

$$Ca(OH)_2 + CO_2 \longrightarrow CaCO_3\downarrow + H_2O$$

Single-stage pH reduction is generally accomplished by the addition of carbon dioxide, although acids have been employed. This reaction, which also represents the second stage of the two-stage method, is as follows:

$$Ca(OH)_2 + 2\,CO_2 \longrightarrow Ca(HCO_3)_2$$

As noted for the other chemicals, the above reactions are merely approximations to the more complex interactions which actually occur in wastewaters.

The lime demand of a given wastewater is a function of the buffer capacity or alkalinity of the wastewater. Figure 5-9 shows this relationship for a number of different wastewaters (6).

5.5 Other Inorganic Chemicals

In addition to aluminum and iron salts and lime a number of other inorganic chemicals have been used in wastewater treatment. Only three are discussed in this section, i.e., soda ash, caustic soda, and carbon dioxide, but others have been and will be employed. Mineral and other acids are prime examples. For information on any of these chemicals, the local supplier or manufacturer should be contacted.

5.5.1 Soda Ash

5.5.1.1 Properties and Availability

Soda ash, Na_2CO_3, is available in two forms. Light soda ash has a bulk density range of 35 to 50 lb/cu ft and a working density of 41 lb/cu ft. Dense soda ash has a density range of 60 to 76 lb/cu ft and a working density of 63 lb/cu ft. The pH of a 1 percent solution of soda ash is 11.2. It is used for pH control and in lime treatment.

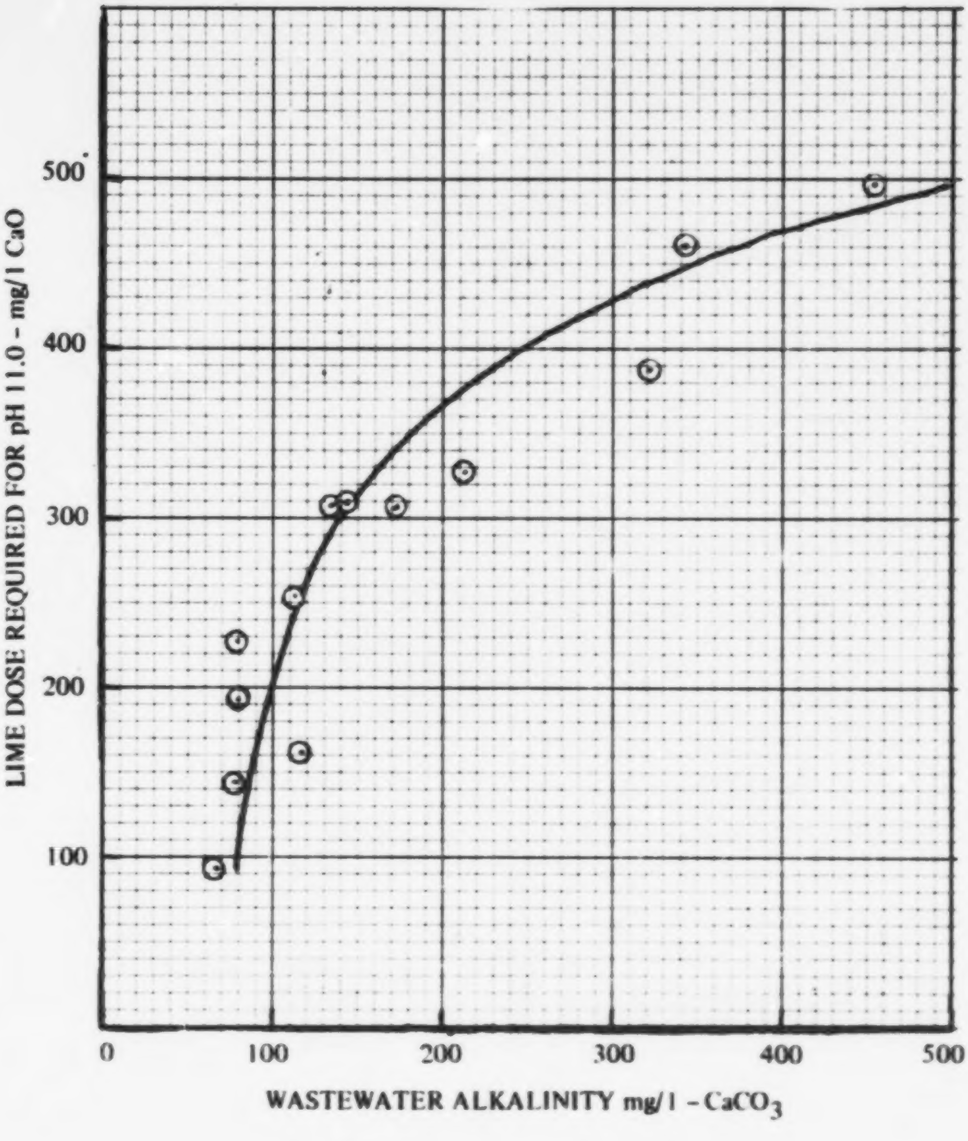


FIGURE 5-9

LIME REQUIREMENT FOR pH ≥ 11.0 AS A FUNCTION OF THE WASTEWATER ALKALINITY

The molecular weight of soda ash is 106. Commercial purity ranges from 98 to greater than 99 percent Na_2CO_3. The viscosities of sodium carbonate solutions are given in Figure 5-10. Soda ash by itself is not particularly corrosive, but in the presence of lime and water, caustic soda is formed which is quite corrosive.

Soda ash is available in bulk by truck, box car and hopper car, and in 100 lb bags from the following partial list of manufacturers.

Location	Manufacturer
CALIFORNIA	
Bartlett	PPG Industries, Inc.
Trona	American Potash and Chemical Corp.
Westend	Stauffer Chemical Co.
GEORGIA	
Brunswick	Allied Chemical Co.
LOUISIANA	
Baton Rouge	Allied Chemical Co.
Lake Charles	Olin Chemicals
MICHIGAN	
Wyandotte	Wyandotte Chemicals Corp.
NEW YORK	
Solvay	Allied Chemical Co.
OHIO	
Barberton	PPG Industries, Inc.
Painesville	Diamond Shamrock Chemical Co.
TEXAS	
Corpus Christi	PPG Industries, Inc.
WEST VIRGINIA	
Moundsville	Allied Chemical Co.
WYOMING	
Green River (3 plants)	Allied Chemical Co. FMC Corp., and Stauffer Chemical Corp.

(COURTESY PPG INDUSTRIES INC., CHEMICAL DIV.)

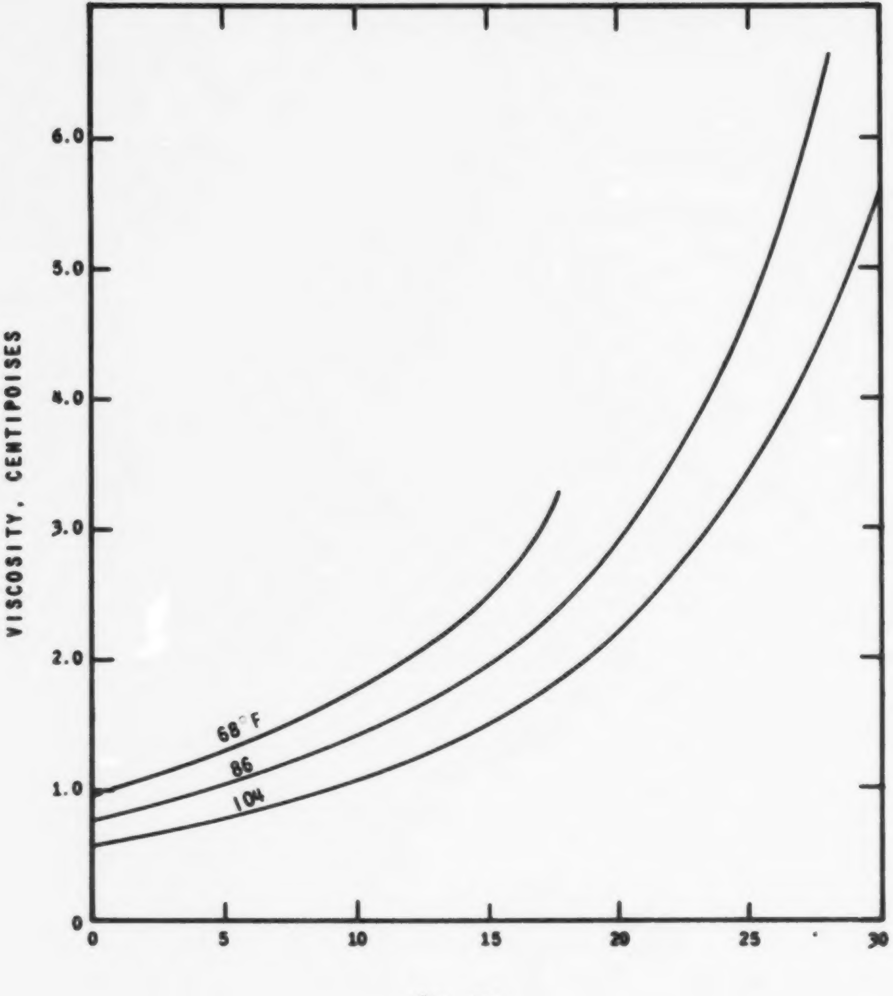


FIGURE 5-10

VISCOSITY OF SODA ASH SOLUTIONS

The current price for soda ash ranges from \$40 to \$50 ton, F.O.B. the point of manufacture, however, prices vary substantially between manufacturers and should be obtained from the closest manufacturers or local distributors. Bagging may add substantially to the cost of the chemical.

5.5.1.2 General Design Considerations

Dense soda ash is generally used in municipal applications because of superior handling characteristics. It has little dust, good flow characteristics, and will not arch in the bin or flood the feeder. It is relatively hard to dissolve and ample dissolver capacity must be provided. Normal practice calls for 0.5 lb of dense soda ash per gal. of water or a 6 percent solution retained for 20 min in the dissolver.

The dust and solution are irritating to the eyes, nose, lungs and skin and therefore general precautions should be observed and the affected areas should be washed promptly with water.

5.5.1.3 Storage

Soda ash is usually stored in steel bins and where pneumatic filling equipment is used, bins should be provided with dust collectors. Bulk and bagged soda ash tend to absorb atmospheric CO_2 and water, forming the less active sodium bicarbonate ($NaHCO_3$). Material recommended for unloading facilities is steel.

5.5.1.4 Feeding Equipment

Feed equipment as described for dry alum is suitable for soda ash. Dissolving of soda ash may be hastened by the use of warm dissolving water. Mechanical or hydraulic jet mixing should be provided in the dissolver.

5.5.1.5 Piping and Accessories

Materials of construction for piping and accessories should be iron, steel, rubber, and plastics.

5.5.1.6 Pacing and Control

Controls as discussed for dry alum apply also to soda ash equipment.

5.5.2 Liquid Caustic Soda

Anhydrous caustic soda ($NaOH$) is available but its use is generally not considered practical in water and wastewater treatment applications. Consequently, only liquid caustic soda is discussed below.

5.5.2.1 Properties and Availability.

Liquid caustic soda is shipped at two concentrations, 50 percent and 73 percent NaOH. The densities of the solutions as shipped are 12.76 lb/gal for the 50 percent solution and 14.18 lb/gal for the 73 percent solution. These solutions contain 6.38 lb/gal NaOH and 10.34 lb/gal. NaOH, respectively. The crystallization temperature is 53°F for the 50 percent solution and 165°F for the 73 percent solution. The molecular weight of NaOH is 40. Viscosities of various caustic soda solutions are presented in Figure 5-11. The pH of a 1 percent solution of caustic soda is 12.9.

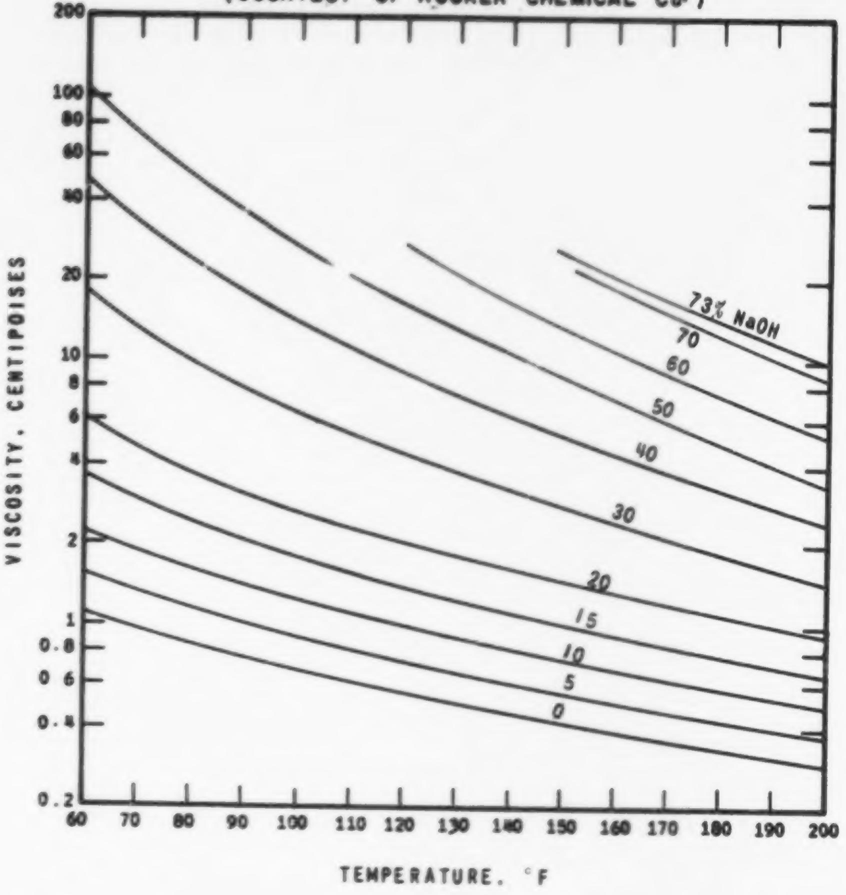
Truck load lots of 1,000 to 4,000 gal are available in the 50 percent concentration only. Both shipping concentrations can be obtained in 8,000, 10,000 and 16,000 gal car load lots. Tank cars can be unloaded through the dome eduction pipe using air pressure or through the bottom valve by gravity or by using air pressure or a pump. Trucks are usually unloaded by gravity or with air pressure or a truck mounted pump.

Major producers of caustic soda and their respective plant locations are listed in Table 5-7. The current price for liquid caustic soda ranges from $76/ton @ 50 percent and $81/ton @ 73 percent, (NaOH), F.O.B. the point of manufacture.

TABLE 5-7

PARTIAL LIST OF CAUSTIC SODA MANUFACTURING PLANTS

Location	Manufacturer
ALABAMA	
Lemoyne (Mobile)	Stauffer
McIntosh	Olin
Muscle Shoals	Diamond Shamrock
CALIFORNIA	
Pittsburg	Dow
DELAWARE	
Delaware City	Diamond Shamrock
GEORGIA	
Augusta	Olin
Brunswick	Allied
KANSAS	
Wichita	Vulcan
KENTUCKY	
Calvert City (2 plants)	Pennwalt, Goodrich



(COURTESY OF HOOKER CHEMICAL Co.)

FIGURE 5-11

VISCOSITY OF CAUSTIC SODA SOLUTIONS

Location	Manufacturer
LOUISIANA	
Baton Rouge	Allied
Geismar	Wyandotte
Lake Charles (2 plants)	PPG, Olin
Plaquemine	Dow
St. Gabriel	Stauffer
Taft	Hooker
MICHIGAN	
Midland	Dow
Montague	Hooker
Wyandotte (2 plants)	Pennwalt, Wyandotte
NEVADA	
Henderson	Stauffer
NEW JERSEY	
Linden	GAF
NEW YORK	
Niagara Falls (3 plants)	Hooker, Olin, Stauffer
Solvay	Allied
NORTH CAROLINA	
Acme	Allied
OHIO	
Barberton	PPG
Cleveland	Harshaw
Painesville	Diamond Shamrock
OREGON	
Portland	Pennwalt
TENNESSEE	
Charleston	Olin
TEXAS	
Corpus Christi	PPG
Deer Park (Houston)	Diamond Shamrock
Freeport	Dow
Port Neches	Jefferson

Location	Manufacturer
VIRGINIA	
Saltville	Olin
WASHINGTON	
Tacoma (2 plants)	Hooker, Pennwalt
WEST VIRGINIA	
Moundsville	Allied
Natrium	PPG
South Charleston	FMC

Manufacturers and addresses

Allied Chemical
Solvay Process Division
40 Rector Street
New York, New York 10006

Diamond Shamrock Chemical Co.
300 Union Commerce Building
Cleveland, Ohio 44115

Dow Chemical Co.
Abbott Road
Midland, Michigan 48640

GAF Corp. Chemical Division
140 West 51st Street
New York, New York 10019

FMC Corporation
Inorganic Chemicals Div.
633 Third Avenue
New York, New York 10017

B.F. Goodrich Chemical Co.
3135 Euclid Avenue
Cleveland, Ohio 44115

Harshaw Chemical Co.
1945 East 97th Street
Cleveland, Ohio 444106

Hooker Chemical Corp.
P.O. Box 344
Niagara Falls, New York 14302

Jefferson Chemical Co., Inc.
3336 Richmond Avenue
Houston, Texas 77006

Olin Corporation
Chemicals Division
745 Fifth Avenue
New York, New York 10022

Pennwalt Corporation
Pennwalt Building
Three Penn Center
Philadelphia, Pa. 19102

PPG Industries, Inc.
Chemical Division
1 Gateway Center
Pittsburgh, Pa. 15222

Stauffer Chemical Co.
Industrial Chemical Div.
299 Park Avenue
New York, New York 10017

Vulcan Materials Co.
Chemicals Division
P.O. Box 545-T
Wichita, Kansas 67201

Wyandotte Chemicals Corp.
Michigan Alkali Division
Wyandotte, Michigan 48192

5.5.2.2 General Design Considerations

Liquid caustic soda is received in bulk shipments, transferred to storage and diluted as necessary for feeding to the points of application. Caustic soda is poisonous and is dangerous to handle. U.S. Department of Transportation Regulations for "White Label" materials must be observed. However, if handled properly caustic soda poses no particular industrial hazard. To avoid accidental spills, all pumps, valves, and lines should be checked regularly for leaks. Workmen should be thoroughly instructed in the precautions related to the handling of caustic soda. The eyes should be protected by goggles at all times when exposure to mist or splashing is possible. Other parts of the body should be protected as necessary to prevent alkali burns. Areas exposed to caustic soda should be washed with copious amounts of water for 15 min to 2 hr. A physician should be called when exposure is severe. Caustic soda taken internally should be diluted with water or milk and then neutralized with dilute vinegar or fruit juice. Vomiting may occur spontaneously but should not be induced except on the advice of a physician.

5.5.2.3 Storage

Liquid caustic soda may be stored at the 50 percent concentration. However, at this solution strength, it crystallizes at 53°F. Therefore, storage tanks must be located indoors or provided with heating and suitable insulation if outdoors. Because of its relatively high crystallization temperature, liquid caustic soda is often dilluted to a concentration of about 20 percent NaOH for storage. A 20 percent solution of NaOH has a crystallization temperature of about -20°F. Recommendations for dilution of both 73 percent and 50 percent solutions should be obtained from the manufacturer, because special considerations are necessary.

Storage tanks for liquid caustic soda should be provided with an air vent for gravity flow. The storage capacity should be equal to 1 1/2 times the largest expected delivery, with an allowance for dilution water, if used, or 2-weeks supply at the anticipated feed rate, whichever is greater. Tanks for storing 50% solution at a temperature between 75°F and

140°F may be constructed of mild steel. Storage temperatures above 140°F require more elaborate materials selection and are not recommended. Caustic soda will tend to pick up iron when stored in steel vessels for extended periods. Subject to temperature and solution strength limitations, rubber, 316 stainless steel, nickel, nickel alloys, or plastics may be used when iron contamination must be avoided.

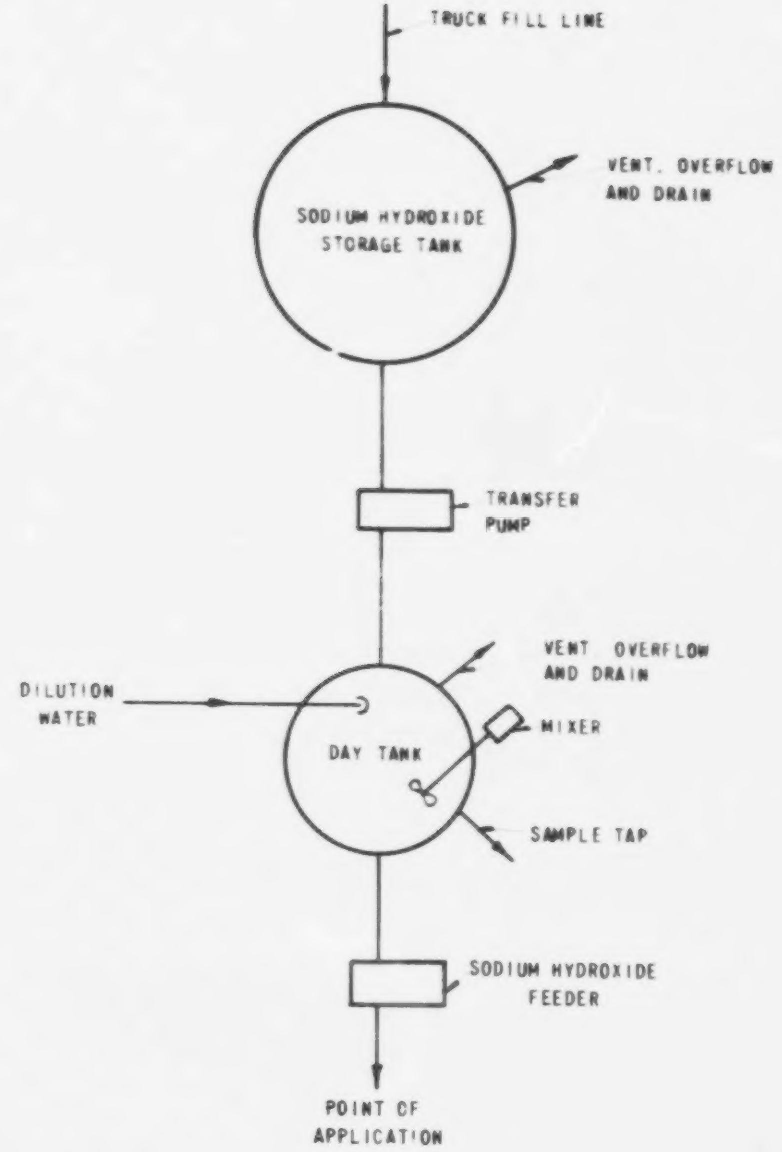
5.5.2.4 Feeding Equipment

Further dilution of liquid caustic soda below the storage strength may be desirable for feeding by volumetric feeders. Feeding systems as described for liquid alum generally apply to caustic soda with appropriate selection of materials of construction. A typical system schematic is shown in Figure 5-12. Feeders will usually include materials such as ductile iron, stainless steels, rubber, and plastics.

5.5.2.5 Piping and Accessories

Transfer lines from the shipping unit to the storage tank should be spiral-wire-bound neoprene or rubber hose, solid steel pipe with swivel joints, or steel hose. Because caustic soda attacks glass, use of glass materials should be avoided. Other miscellaneous materials for use with liquid caustic soda feeding and handling equipment are listed below (7):

Components	Recommended Materials for Use With 50% NaOH Up to 140°F
Rigid Pipe	Standard Weight Black Iron
Flexible Connections	Rigid Pipe with Ells or Swing Joints, Stainless Steel or Rubber Hose
Diluting Tees	Type 304 Stainless Steel
Fittings	Steel
Permanent Joints	Welded or Screwed Fittings
Unions	Screwed Steel
Valves—Non-leaking (Plug)	
Body	Steel
Plug	Type 304 Stainless Steel
Pumps (Centrifugal)	
Body	Steel
Impeller	Ni-Resist
Packing	Blue Asbestos
Storage Tanks	Steel



TRUCK FILL LINE
VENT, OVERFLOW AND DRAIN
SODIUM HYDROXIDE STORAGE TANK
TRANSFER PUMP
VENT OVERFLOW AND DRAIN
DILUTION WATER
MIXER
DAY TANK
SAMPLE TAP
SODIUM HYDROXIDE FEEDER
POINT OF APPLICATION

FIGURE 5-12
TYPICAL CAUSTIC SODA FEED SYSTEM

5.5.2.6 Pacing and Control

Controls as listed for liquid alum also apply to liquid caustic soda equipment.

5.5.3 Carbon Dioxide

5.5.3.1 Properties and Availability

Carbon dioxide, CO_2, is available for use in wastewater treatment plants in gas and liquid form. The molecular weight of CO_2 is 44. Dry CO_2 is not chemically active at normal temperatures and is a non-toxic safe chemical; however, the gas displaces oxygen and adequate ventilation of closed areas should be provided. Solutions of CO_2 in water are very reactive chemically and form carbonic acid. Saturated solutions of CO_2 have a pH of 4.0 at 68°F.

The gas form may be produced on the treatment plant site by scrubbing and compressing the combustion product of lime recalcining furnaces, sludge furnaces, or generators used principally for the production of CO_2 gas only. These generators are usually fired with combustible gases, fuel oil, or coke and have CO_2 yields as shown in Table 5-8 (8).

TABLE 5-8

CO_2 YIELDS OF COMMON FUELS

Fuel	Quantity	CO_2 Yield lb
Natural Gas	1,000 cu ft	115
Coke	1 lb	3
Kerosene	1 gal.	20
Fuel Oil (No. 2)	1 gal.	23
Propane	1,000 cu ft	141
Butane	1,000 cu ft	142

The gas forms, as generated at the plant site, usually have a CO_2 content of between 6 percent and 18 percent depending on the source and efficiency of the producing system.

The liquid form is available from commercial suppliers in 20 to 50 lb cylinders, 10 to 20 ton trucks and 30 to 50 ton rail cars. The commercial liquid form has a minimum CO_2 content of 99.5 percent.

Current prices range from $30 ton for 3,000 tons per year and over, to $68 ton for a quantity of 150 tons year. These prices include an allowance for freight within a 100 mile radius of the point of manufacture. Another $6 ton may be added for each additional 100 miles to the point of destination. Major producers of commercial CO_2 are listed in Table 5-9.

TABLE 5-9

PARTIAL LIST OF CARBON DIOXIDE MANUFACTURING PLANTS

Location	Manufacturer
CALIFORNIA	
Watson (Los Angeles)	Liquid Carbonic
Oakland	Liquid Carbonic
Brea	Airco
Lathrop	Airco
Ventura	Cardox
Taft	Standard Oil
GEORGIA	
Augusta	Liquid Carbonic
ILLINOIS	
Morris	Cardox
Chicago	Airco
INDIANA	
Jeffersonville	Cardox
IOWA	
Clinton	Airco
Ft. Madison	Liquid Carbonic
Ft. Dodge	Liquid Carbonic
Muscatine	Publicker
KANSAS	
Dodge City	Liquid Carbonic
Lawrence	Airco (1972)
Lawrence	Cardox
KENTUCKY	
Doerun (Brandenburg)	Olin
LOUISIANA	
New Orleans	Liquid Carbonic
Luling	Airco
MASSACTUSETTS	
Tewksbury	Liquid Carbonic
MISSISSIPPI	
Yazoo City	Airco
MISSOURI	
Kansas City	Airco
Le May (St. Louis)	Cardox

Location	Manufacturer
NEW JERSEY	
Paulsboro	Olin
Belleville	Liquid Carbonic
Deepwater	Airco
NEW MEXICO	
Bueyeros	SEC
Solana	SEC
Mosquero	SEC
NEW YORK	
Olean	Airco
OHIO	
Toledo	Cardox
Oregon (Toledo)	Liquid Carbonic
Lima	Airco
Huron	Cardox
PENNSYLVANIA	
Philadelphia	Liquid Carbonic
Thermice (Philadelphia)	Publicker
TENNESSEE	
Woodstock (Memphis)	Cardox
TEXAS	
Texas City	Liquid Carbonic
Dallas	Cardox
Dumas	Diamond Shamrock
VIRGINIA	
Hopewell	Airco
Saltville	Olin
WASHINGTON	
Finley	Airco

Manufacturers and Addresses

Cardox Div. of Chemetron Corp.
Dept. TR
840 N. Michigan Avenue
Chicago, Illinois 60611

Airco Industrial Gases Div.of
Air Reduction Co.
575 Mountain Avenue
Murray Hill, N.J. 07974

Olin Corporation
Chemicals Division
745 Fifth Avenue
New York, New York 10022

Publicker Ind., Inc.
Walnut & Thomas
Philadelphia, Pa.

Liquid Carbonic Corporation
Dept. TR
135 S. LaSalle
Chicago, Illinois 60603

SEC
1033 Humble Place
El Paso, Texas 79987

Diamond Shamrock Chemical Co
300 Union Commerce Building
Cleveland, Ohio 44115

Standard Oil Company of California
225 Bush
San Francisco, California 94104

5.5.3.2 General Design Considerations

Recovery of CO_2 from recalcining furnaces or incinerators is the least expensive source, but maintenance of stack gas systems is likely to be extensive because of the corrosive nature of the wet gas and the presence of particulate matter. Scrubber systems are required to clean the stack gas and specially designed gas compressors are necessary to provide the process injection pressure.

Pressure generators and submerged burners require less maintenance because the system pressure is established by compressors or blowers handling dry air or gas. On-site generating units have a limited range of CO_2 production as compared with the liquid storage and feed system, and therefore may require multiple units.

The liquid CO_2 storage and feed system generally includes a temperature-pressure controlled, bulk storage tank, an evaporation unit, and a gas feeder to meter the gas. Solution feeders, similar in construction to chlorinators, may also be used to feed CO_2.

5.5.3.3 Storage

This section applies only to use of commercial liquid CO_2. Liquid system capacities encountered in wastewater treatment usually require on-site bulk storage units. Standard pre-packaged units are available, ranging in size from $^{3}/_{8}$ to 50 tons capacity, and are furnished with temperature-pressure controls to maintain approximately 300 psi at 0°F conditions. The typical package unit contains refrigeration, vaporization, safety and control equipment. The units are well insulated and protected for outdoor location. The gas from the evaporation unit usually passes through two stages of pressure reduction before entering the gas feeder to prevent the formation of dry ice.

5.5.3.4 Feeding Equipment

Feeding systems for the stack gas source of CO_2 consist of simple valving arrangements, for admitting varying quantities of make-up air to the suction side of the constant volume compressors, or for venting excess gas on the compressor discharge. A typical system is described elsewhere (9).

Pressure generators and submerged burners are regulated by valving arrangements on the fuel and air supply. Generation of CO_2 by combustion is usually difficult to control, requires frequent operator attention and demands considerable maintenance over the life of the equipment, when compared with liquid CO_2 systems.

Commercial liquid carbon dioxide is becoming more generally used because of its high purity, the simplicity and range of feeding equipment, ease of control, and smaller, less expensive piping systems. After vaporization, the CO_2 with suitable metering and pressure reduction may be fed directly to the point of application as a gas. However, vacuum operated, solution type gas feeders are often preferred. Such feeders generally include safety devices and operating controls in a compact panel housing, with materials of construction suitable for CO_2 service. Absorption of CO_2 in the injector water supply approaches 100% when a ratio of 1.0 lb of gas to 60 gal of water is maintained.

5.5.3.5 Piping and Accessories

Mild steel piping and accessories are suitable for use with cool, dry, carbon dioxide. Hot, moist gases, however, require the use of type 316 stainless steel or plastic materials. Plastics or FRP pipe are generally used for solution piping and diffusers. Diffusers should be submerged at least 8 ft, and preferably deeper, to assure complete absorption of the gas.

5.5.3.6 Pacing and Control

Standard instrument signals and control components can be used to pace or control carbon dioxide feed systems.

Using stack gas as the source of CO_2, the feed rate can be controlled by proper selection and operation of compressors, by manual control of vent or bleed valves, or by automatic control of such valves by a pH meter-controller system.

In commercial CO_2 feed systems, solution feeders may function as controllers and can be paced by instrument signals from pH monitors and plant flow meters.

In feeding commercial CO_2 directly to the point of application as a gas, a differential pressure transmitter and a control valve may function as the primary elements of a control system. Standard instrument signals may be used to pace or control the rate of CO_2 feed.

CO_2 generators are difficult to pace or control other than by manual or automatic operation of vent or bleed valves that waste a portion of the produced gas according to the plant requirements.

5.6. Polymers

Polymeric flocculants are high molecular weight organic chains with ionic or other functional groups incorporated at intervals along the chains. Because these compounds have characteristics of both polymers and electrolytes, they are frequently called polyelectrolytes. They may be of natural or synthetic origin.

All synthetic polyelectrolytes can be classified on the basis of the type of charge on the polymer chain. Thus polymers possessing negative charges are called anionic while those carrying positive charges are cationic. Certain compounds carry no electrical charge and are called nonionic polyelectrolytes.

Because of the great variety of monomers available as starting material and the additional variety that can be obtained by varying the molecular weight, charge density and ionizable groups, it is not surprising that a great assortment of polyelectrolytes are available to the wastewater plant operator. A partial listing of manufacturers is shown in Table 5-10. This list is based mainly on three major sources (10) (11) (12) and does not purport to be a complete list.

Extensive use of any specific polymer as a flocculant is of necessity determined by the size, density and ionic charge of the colloids to be coagulated. As other factors need to be considered, i.e. coagulants used, pH of the system, techniques and equipment for dissolution of the polyelectroyte, etc., it is mandatory that extensive jar testing be performed to determine the specific polymer that will perform its function most efficiently. These results should be verified by plant-scale testing.

5.6.1 Dry Polymers

5.6.1.1 Properties and Availability

Types of polymers vary widely in characteristics. Manufacturers should be consulted for properties, availability, and cost of the polymer being considered. References are available that indicate the types and characteristics of polymers available (10) (11) (12). Bulk shipments are generally not desirable. Polymers are available in a variety of container or package sizes.

TABLE 5-10

PARTIAL LIST OF POLYMER SOURCES AND TRADE NAMES

Source	Trade Name (s)
Allied Colloids, Inc. One Robinson Lane Ridgewood, N.J. 07450	Percol
Allstate Chemical Co. Box 3040 Euclid, Ohio 44117	Allstate

TABLE 5-10 (continued)

Source	Trade Name (s)
Allyn Chemical Co 2224 Fairhill Rd Cleveland, Ohio 44106	Claron
American Cyanamid Co Berdan Ave. Wayne, N.J. 07470	Superfloc Magnifloc
Atlas Chemical Ind., Inc. Wilmington, Dela. 19899	Sorbo Atlasep
Berdell Industries 28-01 Thomson Ave. Long Island City, NY 11101	Berdell
Betz Laboratories, Inc. Somerton Red. Trevose, Pa.19047	Betz Polyfloc
Bond Chemicals, Inc. 1500 Brookpark Rd. Cleveland, Ohio 44109	Bondfloc
Brennan Chemical Co. 704 N. First St. St. Louis, Mo. 63102	Brenco
The Burtonite Company Nutley, N.J. 07110	Burtonite
Calgon Corporation P.O. Box 1346 Pittsburgh, Pa. 15222	Cat-Floc
Commercial Chemical 11 Paterson Ave. Midland Park, N.J. 07432	Speedifloc

TABLE 5-10 (continued)

Source	Trade Name (s)
Dearborn Chemical Div. W.R. Grace & Co. Merchandise Mart Plaza Chicago, Ill. 60654	Aquafloc
Dow Chemical USA Barstow Building 2020 Dow Center Midland, MI. 48640	Dowell PEI Purifloc Separan XD
Drew Chemical Corp. 701 Jefferson Rd. Parisippany, N.J. 07054	Drewfloc Amerfloc
Du Bois Chemicals Div. W.R. Grace & Co. 3630 E. Kemper Rd. Sharonville, Ohio 45241	Flocculite
E.I. DuPont de Nemours & Co. Eastern Laboratory Gibbstown, N.J. 08027	Du Pont
Environmental Pollution Investigation & Control, Inc. 9221 Bond St. Overland Park, KS. 66214	Dynafloc
Fabcon International 1275 Columbus Ave. San Francisco, Calif. 94133	Zuclar Fabcon
Henry W. Fink & Co. 6900 Silverton Avenue Cincinnati, Ohio 45236	Kleer-Floc
Gamlen Sybron 321 Victory Avenue S. San Francisco, Calif. 94080	Gamafloc Gamlose Gamlen

TABLE 5-10 (continued)

Source	Trade Name(s)
Garrett-Callahan 111 Rollins Rd. Millbrae, Calif. 94030	Garrett-Callahan
General Mills Chemicals 4620 N. 77th Street Minneapolis, Min. 55435	Supercol Guartec
Hercules, Inc. 910 Market St. Wilmington, Dela. 19899	Hercofloc
Frank Herzl Corp. 299 Madison Avenue New York, N.Y. 10017	Perfectamyl
ICI America, Inc. Wilmington, Dela. 19899	Atlasep
Illinois Water Treatment Co. 840 Cedar St. Rockford, Ill. 61102	Illco
Kelco Company 8225 Aero Dr. San Diego, Calif. 92123	Kelgin Kelcosol
Key Chemicals 4346 Tacony Philadelphia, Pa. 19124	Key-Floc
Metalene Chemical Co. Bedford, Ohio 44014	Metalene
The Mogul Corporation 20600 Chagrin Blvd. Cleveland, Ohio 44122	Mogul
Nalco Chemical Co. 6216 W. 66th Street Chicago, Ill. 60638	Nalcolyte

TABLE 5-10 (continued)

Source	Trade Name (s)
Narvon Mining & Chemical Co. Keller Ave. & Fruitville Pike Lancaster, Pa. 17604	Sink-Floc Zeta-Floc
National Starch & Chemical Corp. 1700 W. Frront St. Plainfield, N.J. 07063	Floc-Aid Natron
O'Brien Industries, Inc. 95 Dorsa Avenue Livingston, N.J. 07039	O'B Floc
Oxford Chemical Div. Consolidated Foods Corp. P.O. Box 80202 Atlanta, Ga. 30341	Oxford-Hydro-Floc
Reichhold Chemicals, Inc. RCI Building White Plains, N.Y. 10602	Aquarid
Standard Brands Chem. Ind., Inc. P.O. Drawer K Dover, Dela. 19901	Tychen
A.E. Staley Mfg. Co. P.O. Box 151 Decatur, Ill. 62525	Hamaco
Stein, Hall & Co., Inc. 605 Third Avenue New York, N.Y. 10016	Hallmark Jaguar Polyhall
Swift & Company Oakbrook, Ill. 60521	Swift
James Varley & Sons, Inc. 1200 Switzer Ave. St. Louis, Mo. 63147	Varco-Floc
W.E. Zimmie, Inc. 810 Sharon Dr. Westlake, Ohio 44145	Zimmite

5.6.1.2 General Design Considerations

Dry Polymer and water must be blended and mixed to obtain a recommended solution for efficient action. Solution concentrations vary from fractions of a percent up. Preparation of the stock solution involves wetting of the dry material and usually an aging period prior to application. Solutions can be very viscous, and close attention should be paid to piping size and length and pump selections. Metered solution is usually diluted just prior to injection to the process to obtain better dispersion at the point of application.

5.6.1.3 Storage

General practice for storage of bagged dry chemicals should be observed. The bags should be stored in a dry, cool, low humidity area and used in proper rotation, i.e., first in, first out.

Solutions are generally stored in type 316 stainless steel, FRP, or plastic lined tanks.

5.6.1.4 Feed Equipment

Two types of systems are frequently combined to feed polymers. The solution preparation system includes a manual or automatic blending system with the polymer dispensed by hand or by a dry feeder to a wetting jet and then to a mixing-aging tank at a controlled ratio. The aged polymer is transported to a holding tank where metering pumps or rotodip feeders dispense the polymer to the process. A schematic of such a system is shown by Figure 5-13. It is generally advisable to keep the holding or storage time of polymer solutions to a minimum, 1 to 3 days or less, to prevent deterioration of the product.

5.6.1.5 Piping and Accessories

Selection must be made after determination of the polymer, however, type 316 stainless steel or plastics are generally used.

5.6.1.6 Pacing and Controls

Controls as listed for liquid alum apply to the control of liquid dispersing feeders.

The solution preparation system may be an automatic batching system, as shown by the schematic on Figure 5-14, that fills the holding tank with aged polymer as required by level probes. Such a system is usually provided only at large plants. Prepackaged solution preparation units are available, but have a limited capacity.

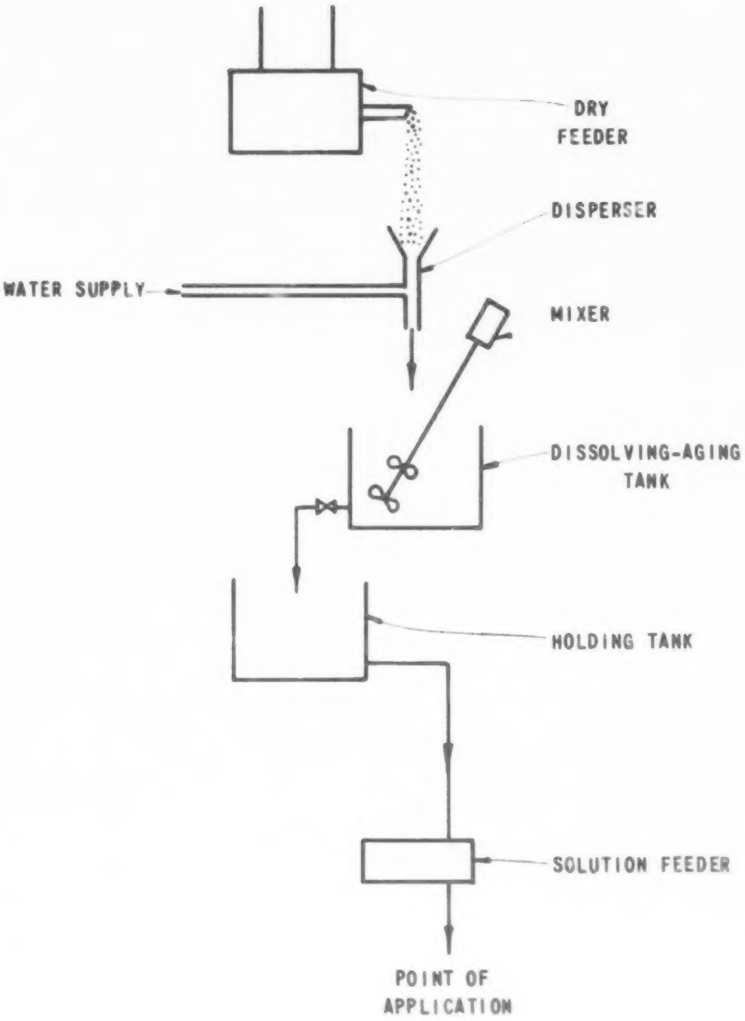
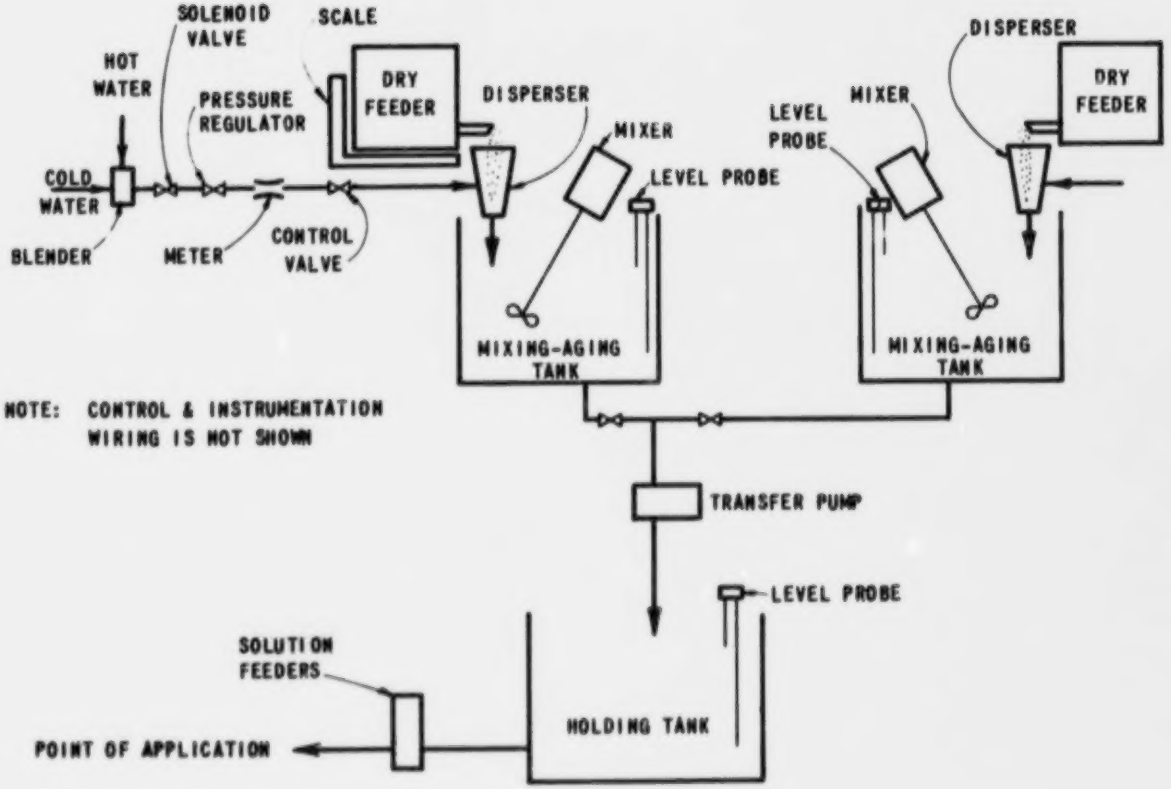

DRY FEEDER
DISPERSER
WATER SUPPLY
MIXER
DISSOLVING-AGING TANK
HOLDING TANK
SOLUTION FEEDER
POINT OF APPLICATION

FIGURE 5-13
TYPICAL SCHEMATIC OF A DRY POLYMER
FEED SYSTEM


SOLENOID VALVE
SCALE
HOT WATER
PRESSURE REGULATOR
DRY FEEDER
DISPERSER
MIXER
LEVEL PROBE
COLD WATER
BLENDER
METER
CONTROL VALVE
MIXING-AGING TANK
DISPERSER
DRY FEEDER
LEVEL PROBE
MIXER
MIXING-AGING TANK
TRANSFER PUMP
LEVEL PROBE
SOLUTION FEEDERS
HOLDING TANK
POINT OF APPLICATION

NOTE: CONTROL & INSTRUMENTATION WIRING IS NOT SHOWN

FIGURE 5-14
TYPICAL AUTOMATIC DRY POLYMER FEED SYSTEM

5.6.2 Liquid Polymers

5.6.2.1 Properties and Availability

As with dry polymers, there is a wide variety of products, and manufacturers should be consulted for specific information.

5.6.2.2 General Design Considerations

Liquid systems differ from the dry systems only in the equipment used to blend the polymer with water to prepare the solution. Liquid solution preparation is usually a hand batching operation with manual filling of a mixing-aging tank with water and polymer.

5.6.2.3. Feed Equipment

Liquid Polymers need no aging and simple dilution is the only requirement for feeding. The dosage of liquid polymers may be accurately controlled by metering pumps or rotodip feeders.

The balance of the process is generally the same as described for dry polymers.

5.7 Chemical Feeders

Chemical feed systems must be flexibly designed to provide for a high degree of reliability in light of the many contingencies which may affect their operation. Thorough waste characterization in terms of flow extremes and chemical requirements should precede the design of the chemical feed system. The design of the chemical feed system must take into account the form of each chemical desired for feeding, the particular physical and chemical characteristics of the chemical, maximum waste flows and the reliability of the feeding devices.

In suspended and colloidal solids removal from wastewaters the chemicals employed are generally in liquid or solid form. Those in solid form are generally converted to solution or slurry form prior to introduction to the wastewater stream; however, some chemicals are fed in a dry form. In any case, some type of solids feeder is usually required. This type of feeder has numerous different forms due to wide ranges in chemical characteristics, feed rates and degree of accuracy required. Liquid feeding is somewhat more restrictive, depending mainly on liquid volume and viscosity.

The capacity of a chemical feed system is an important consideration in both storage and feeding. Storage capacity design must take into account the advantage of quantity purchase versus the disadvantage of construction cost and chemical deterioration with time (13). Potential delivery delays and chemical use rates are necessary factors in the total picture. Storage tanks or bins for solid chemicals must be designed with proper consideration of the angle of repose of the chemical and its necessary environmental requirements, such as temperature and humidity. Size and slope of feeding lines are important along with their materials of construction with respect to the corrosiveness of the chemicals.

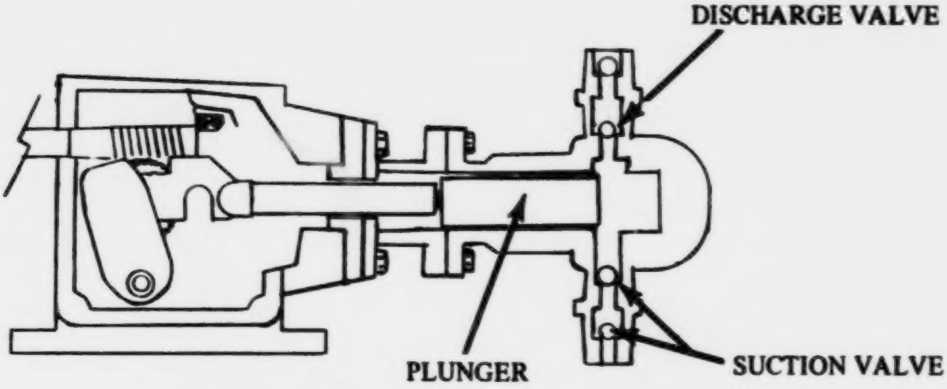
Chemical feeders must accommodate the minimum and maximum feeding rates required. Baker (13) indicates that manually controlled feeders have a common range of 20:1, but this range can be increased to about 100:1 with dual control systems. Chemical feeder control can be manual, automatically proportioned to flow, dependent on some form of process feedback or a combination of any two of these. More sophisticated control systems are feasible if proper sensors are available. If manual control systems are specified with the possibility of future automation, the feeders selected should be amenable to this conversion with a minimum of expense. An example would be a feeder with an external motor which could easily be replaced with a variable speed motor or drive when automation is installed (13). Standby or backup units should be included for each type of feeder used. Reliability calculations will be necessary in larger plants with a greater multiplicity of these units. Points of chemical addition and piping to them should be capable of handling all possible changes in dosing patterns in order to have proper flexibility of operation. Designed flexibility in hoppers, tanks, chemical feeders and solution lines is the key to maximum benefits at least cost (14).

Liquid feeders are generally in the form of metering pumps or orifices. Usually these metering pumps are of the positive-displacement variety, plunger or diaphragm type. The choice of liquid feeder is highly dependent on the viscosity, corrosivity, solubility, suction and discharge heads, and internal pressure-relief requirements (10). Some examples are shown in Figure 5-15. In some cases control valves and rotameters may be all that is required. In other cases, such as lime slurry feeding, centrifugal pumps with open impellers are used with appropriate controls (9). More complete descriptions of liquid feeder requirements can be found in the literature and elsewhere (14).

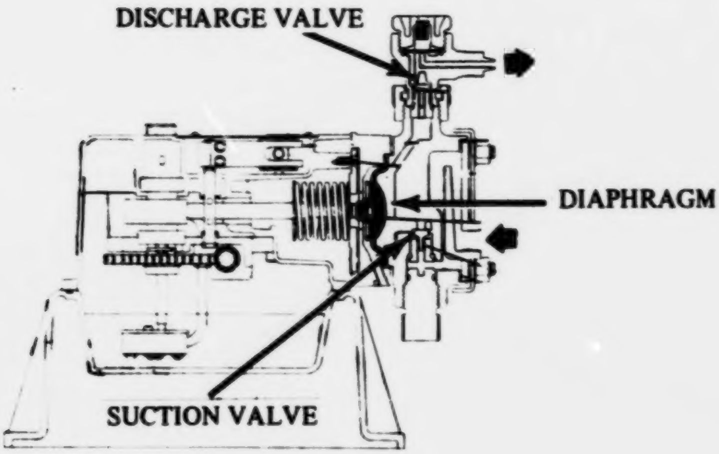
Solids characteristics vary to a great degree and the choice of feeder must be considered carefully, particularly in the smaller-sized facility where a single feeder may be used for more than one chemical. Generally, provisions should be made to keep all chemicals cool and dry. Dryness is very important, as hygroscopic (water absorbing) chemicals may become lumpy, viscous or even rock hard; other chemicals with less affinity for water may become sticky from moisture on the particulate surfaces, causing increased arching in hoppers. In either case, moisture will affect the density of the chemical and may result in under-feed. Dust removal equipment should be used at shoveling locations, bucket elevators, hoppers and feeders for neatness, corrosion prevention and safety reasons. Collected chemical dust may often be used.

The simplest method for feeding solid chemicals is by hand. Chemicals may be preweighed or simply shoveled or poured by the bagful into a dissolving tank. This method is of economic necessity limited to very small operations, or to chemicals used in very weak solutions.

Because of the many factors, such as moisture content, different grades and compressibility, which can affect chemical density (weight to volume ratio), volumetric feeding of solids is normally restricted to smaller plants, specific types of chemicals which are reliably constant in composition and low rates of feed. Within these restrictions several volumetric types are


DISCHARGE VALVE
PLUNGER
SUCTION VALVE

PLUNGER PUMP (Courtesy of Wallace & Tiernan)


DISCHARGE VALVE
DIAPHRAGM
SUCTION VALVE

DIAPHRAGM PUMP (Courtesy of Wallace & Tiernan)

FIGURE 5-15
POSITIVE DISPLACEMENT PUMPS

available. Accuracy of feed is usually limited to ± 2 percent by weight but may be as high as ±15 percent.

One type of volumetric dry feeder uses a continuous belt of specific width moving from under the hopper to the dissolving tank. A mechanical gate mechanism regulates the depth of material on the belt, and the rate of feed is governed by the speed of the belt and or the height of the gate opening. The hopper normally is equipped with a vibratory mechanism to reduce arching. This type of feeder is not suited for easily fluidized materials.

Another type employs a screw or helix from the bottom of the hopper through a tube opening slightly larger than the diameter of the screw or helix. Rate of feed is governed by the speed of screw or helix rotation. Some screw-type designs are self-cleaning, while others are subject to clogging. Figure 5-16 shows a typical screw-feeder.

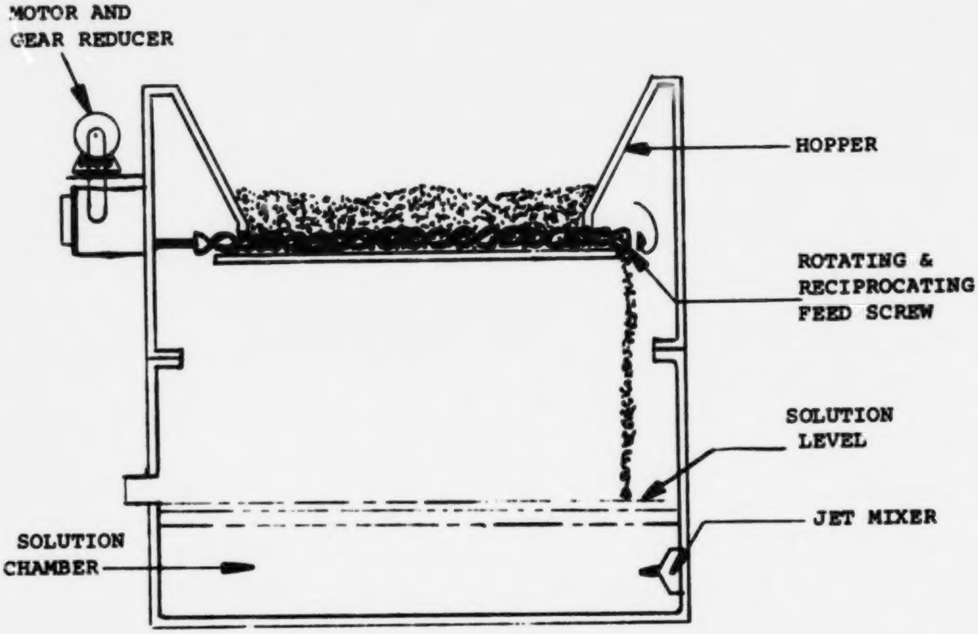
Most remaining types of volumetric feeders generally fall into the positive-displacement category. All designs of this type incorporate some form of moving cavity of a specific or variable size. In operation, the chemical falls by gravity into the cavity and is more or less fully enclosed and separated from the hopper's feed. The size of the cavity, and the rate at which the cavity moves and is discharged, governs the amount of material fed. The positive control of the chemical may place a low limit on rates of feed. One unique design is the progressive cavity metering pump, a non-reciprocating type. Positive-displacement feeders often utilize air injection to improve the flow of the material. Some examples of positive-displacement units are illustrated in Figure 5-17.

The basic drawback of volumetric feeder design, i.e., its inability to compensate for changes in the density of materials, is overcome by modifying the volumetric design to include a gravimetric or loss-in-weight controller. This modification allows for weighing of the material as it is fed. The beam balance type measures the actual mass of material. This is, considerably more accurate, particularly over a long period of time, than the less common spring-loaded gravimetric designs. Gravimetric feeders are used where feed accuracy of about 1 % is required for economy, as in large scale operations and for materials which are used in small, precise quantities. It should be noted, however, that even gravimetric feeders cannot compensate for weight added to the chemical by excess moisture. Many volumetric feeders may be converted to loss-in-weight function by placing the entire feeder on a platform scale which is tared to neutralize the weight of the feeder.

Good housekeeping and need for accurate feed rates dictate that the gravimetric feeder be shut down and thoroughly cleaned on a regular basis. Although many of these feeders have automatic or semi-automatic devices which compensate to some degree for accumulated solids on the weighing mechanism, accuracy is affected, particularly on humid days when hygroscopic materials are fed. In some cases, built-up chemicals can actually jam the equipment.

No discussion of feeders is complete without at least passing reference to dissolvers, as any metered material must be mixed with water to provide a chemical solution of desired strength. Most feeders, regardless of type, discharge their material to a small dissolving

FIGURE 5-16
SCREW FEEDER


MOTOR AND
GEAR REDUCER
HOPPER
ROTATING &
RECIPROCATING
FEED SCREW
SOLUTION
LEVEL
JET MIXER
SOLUTION
CHAMBER

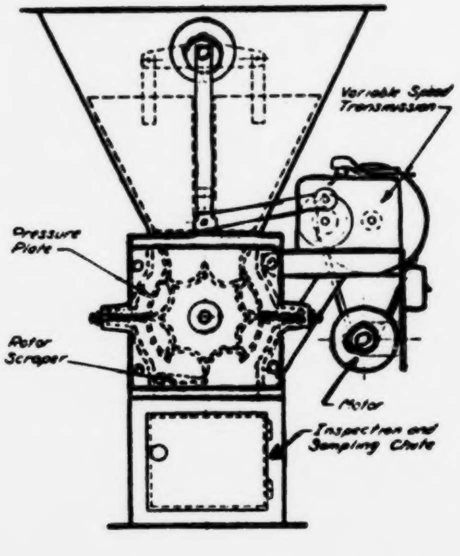

Variable Speed
Transmission
Pressure
Plate
Rotor
Scraper
Motor
Inspection and
Sampling Chute

FIGURE 5-17
POSITIVE DISPLACEMENT SOLID
FEEDER—ROTARY (15)

tank which is equipped with a nozzle system and/or mechanical agitator depending on the solubility of the chemical being fed. Solid materials, such as polyelectrolytes, may be carefully spread into a vortex spray or washdown jet of water immediately before entering the dissolver. It is essential that the surface of each particle become thoroughly wetted before entering the feed tank to ensure accurate dispersal and to avoid clumping, settling or floating.

A dissolver for a dry chemical feeder is unlike a chemical feeding mechanism, which by simple adjustment and change of speed can vary its output tenfold. The dissolver must be designed for the job to be done. A dissolver suitable for a rate of 10 lb/hr may not be suitable for dissolving at a rate of 100 lb/hr. As a general rule, dissolvers may be oversized, but dissolvers for commercial ferric sulfate or lime slakers do not perform well if greatly oversized.

It is essential that specifications for dry chemical feeders include specifications on dissolver capacity. A number of factors need to be considered in designing dissolvers of proper capacity. These include detention times and water requirements, as well as other factors specific to individual chemicals.

The capacity of a dissolver is based on detention time, which is directly related to the wettability or rate of solution of the chemical. Therefore, the dissolver must be large enough to provide the necessary detention for both the chemical and the water at the maximum rate of feed. At lower rates of feed, the strength of solution or suspension leaving the dissolver will be less, but the detention time will be approximately the same unless the water supply to the dissolver is reduced. When the water supply to any dissolver is controlled for the purpose of forming a constant strength solution, mixing within the dissolver must be accomplished by mechanical means, because sufficient power will not be available from the mixing jets at low rates of flow. Hot water dissolvers are also available in order to mimimize the required tankage.

The foregoing descriptions give some indication of the wide variety of materials which may be handled. Because of this variety, a modern facility may contain any number of a variety of feeders with combined or multiple materials capability. Ancillary equipment to the feeder also varies according to the material to be handled. Liquid feeders encompass a limited number of design principles which account for density and viscosity ranges. Solids feeders, relatively speaking, vary considerably due to the wide range of physical and chemical characteristics, feed rates and the degree of precision and repeatability required.

Table 5-11 describes several types of chemical feeders commonly used in wastewater treatment.

TABLE 5-11
TYPES OF CHEMICAL FEEDERS

Type of Feeder	Use	Limitations		
		General	Capacity cu ft/hr	Range
Dry feeder:				
Volumetric:				
Oscillating plate	Any material, granules or powder.		0.01 to 35	40 to 1
Oscillating throat (universal)	Any material, any particle size.		0.02 to 100	40 to 1
Rotating disc	Most materials including NaF, granules or powder.	Use disc unloader for arching.	0.01 to 1.0	20 to 1
Rotating cylinder (star)	Any material, granules or powder.		8 to 2,000 or 7.2 to 300	10 to 1 or 100 to 1
Screw	Dry, free flowing material, powder or granular.		0.05 to 18	20 to 1
Ribbon	Dry, free flowing material, powder, granular, or lumps.		0.002 to 0.16	10 to 1
Belt	Dry, free flowing material up to 1½-inch size, powder or granular.		0.1 to 3,000	10 to 1 or 100 to 1
Gravimetric:				
Continuous - belt and scale	Dry, free flowing, granular material, or floodable material.	Use hopper agitator to maintain constant density.	0.02 to 2	100 to 1
Loss in weight	Most materials, powder, granular or lumps.		0.02 to 80	100 to 1
Solution feeder:				
Nonpositive displacement:				
Decanter (lowering pipe)	Most solutions or light slurries		0.01 to 10	100 to 1
Orifice	Most solutions	No slurries	0.16 to 5	10 to 1
Rotameter (calibrated valve)	Clear solutions	No slurries	0.005 to 0.16 or 0.01 to 20	10 to 1
Loss in weight (tank with control valve).	Most solutions	No slurries	0.002 to 0.20	30 to 1
Positive displacement:				
Rotating dipper	Most solutions or slurries		0.1 to 30	100 to 1
Proportioning pump:				
Diaphragm	Most solutions. Special unit for 5% slurries.[1]		0.004 to 0.15	100 to 1
Piston	Most solutions, light slurries.		0.01 to 170	20 to 1
Gas feeders:				
Solution feed	Chlorine		8000 lb/day max	20 to 1
	Ammonia		2000 lb/day max	20 to 1
	Sulfur dioxide		7600 lb/day max	20 to 1
	Carbon dioxide		6000 lb/day max	20 to 1
Direct feed	Chlorine		300 lb/day max	10 to 1
	Ammonia		120 lb/day max	7 to 1
	Carbon dioxide		10,000 lb/day max	20 to 1

[1] Use special heads and valves for slurries.

5.8 References

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CHAPTER 6

CHEMICAL MIXING, FLOCCULATION AND SOLIDS-CONTACT PROCESSES

6.1 Introduction

Chemical mixing and flocculation or solids-contact are important mechanical steps in the overall coagulation process described in Chapter 4. Application of the processes to wastewater generally follows standard practices and employs basic equipment used for years in the water-treatment field. Chemical mixing thoroughly disperses coagulants or their hydrolysis products so the maximum possible portion of influent colloidal and fine supracolloidal solids are absorbed and destabilized. Flocculation or solids contact processes increase the natural rate of contacts between particles. This makes it possible, within reasonable detention periods, for destabilized colloidal and fine supracolloidal solids to aggregate into particles large enough for effective separation by gravity processes or media filtration.

All the processes discussed in this chapter depend on fluid shear for coagulant dispersal and for promoting particle contacts. Shear is most commonly introduced by mechanical mixing equipment. In certain solids-contact processes shear results from fluid passage upward through a blanket of previously settled particles. Some designs have utilized shear resulting from energy losses in pumps or at ports and baffles.

Numerous theoretical descriptions of the flocculation process have been developed (1) (2) (3) (4) (5) (6). Almost all relate to experience in water treatment but can be applied to wastewater coagulation with proper attention to significant differences in the nature of solids.

All theoretical approaches recognize the importance of time (t in sec.) and velocity gradient ($\overline{G}$, a measure of shear intensity in fps ft or sec^{-1}) as controlling parameters in determining performance of mixing and flocculation processes. It should be noted that chemical mixing and flocculation differ only in intensity and duration and that some aggregation takes place in the chemical mixing stage.

In addition to velocity gradient and time, expressions for the rate of aggregation in flocculation or solids contact processes involve parameters reflecting the total volume and the size and number of floc particles. When destabilized, particles in the fine colloidal range rapidly aggregate under natural conditions to form small flocs of fine supracolloidal size, about 1 micron diameter (5), often termed *primary* sized particles. In developing mathematical relations this is generally the assumed initial size of particles to be further aggregated.

The rate of aggregation is commonly taken as a function of the dimensionless product $\overline{G}Ct$ where C is the ratio of the volume of floc to total volume of suspension and $\overline{G}$ and t are as defined above.

The floc volume concentration resulting from a given coagulant dosage depends, among other things, on the amount of water entrained in the floc. Hudson (7) and Camp (1) have shown that more water is entrained and higher floc volumes result when flocculation takes place at lower values of $\overline{G}$.

The value of C may be increased greatly by recirculation of settled solids. This is used in certain types of solids contact reactors (Section 6.4) and has been applied at Lake Tahoe as part of a conventional coagulation system with separate rapid mix, flocculation and sedimentation basins (8).

Design of rapid mix and flocculation units generally involves the choice of detention and $\overline{G}$ value and selection of configurations, of mixing equipment, tanks, piping, etc. Unless the designer provides for direct control of floc volume concentration through solids recirculation, operating values of this parameter are determined indirectly through the chemical dosage and choice of $\overline{G}$ value and detention. Special attention should be given to avoiding excessive localized shear and reducing short circuiting. Pretreatment should assure that wastewater is free of debris (rags, sticks, etc.) which could damage mixing equipment. Special considerations in design of solids-contact units are presented in Section 6.4.

$\overline{G}$ represents the root mean square velocity gradient (fps/ft) over the mixing basin. For mechanically-stirred basins it can be calculated from the relation:

$$\overline{G} = \left(\frac{P}{Vu}\right)^{\frac{1}{2}}$$

Where: P = power applied to stirring (ft-lb/sec = HP x 550)
V = reactor volume (cu ft)
u = viscosity of fluid (lb-sec/sq ft)

Viscosity varies with temperature as follows:

T °C	u lb-sec/sq ft
1	0.361×10^{-4}
5	0.316×10^{-4}
10	0.273×10^{-4}
15	0.239×10^{-4}
20	0.210×10^{-4}
25	0.187×10^{-4}
30	0.167×10^{-4}

Formulas for calculating $\overline{G}$ from head losses in baffled basins or in conduits are given by Camp (9).

6.2 Chemical Mixing

Chemical mixing facilities should be designed to provide a thorough and complete dispersal of chemical throughout the wastewater being treated to insure uniform exposure to pollutants which are to be removed.

The intensity and duration of mixing of coagulants with wastewater must be controlled to avoid overmixing or undermixing.

Overmixing excessively disperses newly-formed floc and may rupture existing wastewater solids. Excessive floc dispersal retards effective flocculation and may significantly increase the flocculation period needed to obtain good settling properties. The rupture of incoming wastewater solids may result in less efficient removals of pollutants associated with those solids (2) (4).

Undermixing inadequately disperses coagulants resulting in uneven dosing. This in turn may reduce efficiency of solids removal while requiring unnecessarily high coagulant dosages.

In water treatment practice several types of chemical mixing units have been used. These include high-speed mixers, in-line blenders and pumps, and baffled mixing compartments or static in-line mixers (baffled piping sections). The high-speed mixer, as shown in Figure 6-1, has been the most common choice for water treatment. Designs usually call for a 10 to 30 second detention time and approximately 300 fps/ft velocity gradiant (10). Hudson and Wolfner (11) recommend variable-speed mixers to allow for varying requirements for optimum mixing. In solids-contact reactors the $\overline{G}$ values in the immediate mixing zone approximate those for high-speed mixing (See section 6.4).

High speed mixers designed on the above basis should be equally satisfactory for wastewater applications. Culp, et al, (12) recommend providing two parallel units with a somewhat larger detention: 2 minutes at total design flow with both units. It has been questioned, however, whether in-line blenders (with $\overline{G}$ values as high as 5000 fps/ft) should be used for wastewater in view of the possibility of rupturing organic solids (4). Where flows must be pumped just prior to coagulation, addition of chemicals at the pumps is feasible. The pump selection should take into account possible effects on organic solids of shear in centrifugal units. Where problems are anticipated, lower speed units such as screw pumps should be used. Baffled compartments or in-line static mixing devices are limited in their effectiveness as chemical mixing devices whether in water or wastewater treatment because:

1. Head losses of up to 3 ft are required.
2. $\overline{G}$ cannot be changed to meet requirements, but rather is a function of flow rate through the units.

In mineral addition to biological wastewater treatment systems, coagulants may be added directly to mixed biological reactors such as aeration tanks or rotating biological con-

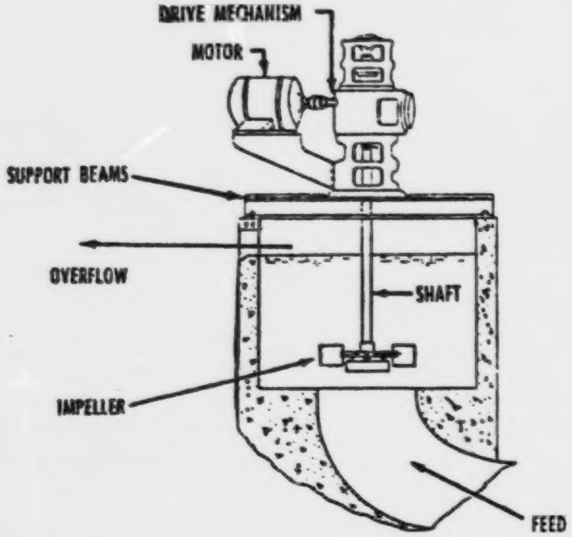
DRIVE MECHANISM
MOTOR
SUPPORT BEAMS
OVERFLOW
SHAFT
IMPELLER
FEED

FIGURE 6-1
IMPELLER MIXER

tactors.

Based on typical power inputs per unit tank volume, mechanical and diffused aeration equipment and rotating fixed-film biological contactors produce average shear intensities generally in the range suitable for chemical mixing. Parker (13) indicated that an analysis of data for 14 activated sludge plants revealed that $\bar{G}$ ranged from 88 to 220 fps/ft with an average of 136 fps/ft. Localized maximum shear intensities vary widely depending on speed of rotating equipment or on bubble size for diffused aeration. Camp (9) presented bases for relating localized maximum shear intensities to bubble size in diffused aeration. For fine bubble diffusion (1.5 mm bubbles) maximum intensity reaches 1500 fps/ft with higher values for coarse bubbles.

No similar development has been located for rotating mechanical aerating equipment, but it appears that maximum localized shears range from little more than the basin mean value for large, low-speed devices such as rotating biological contactors, to perhaps as high as 50 times the mean for high speed (1800 rpm) mechanical aerators. Questions have been raised about detrimental effects of high speed aerators on settling of activated sludge.

When using polymers, manufacturer's recommendations should be sought on the mixing conditions which optimize their effectiveness, and these should be supplemented by jar tests, if possible. When coagulant aids are employed, provisions for multiple addition points should be made at the rapid mixing basin and in the flocculator to optimize the performance of the coagulant.

6.3 Flocculation

The proper measure of flocculation effectiveness is the performance of subsequent solids separation units in terms of both effluent quality and operating requirements, such as filter backwash frequency. Effluent quality depends greatly on the reduction of residual primary size particles during flocculation, while operating requirements relate more to the floc volume applied to separation units.

For water treatment using alum or iron coagulants and flow-through flocculation (as opposed to solids-contact units) traditional designs have been based on $\bar{G}$ of up to 100 fps/ft and $\bar{G}t$ values of 0.3 to 1.5×10^5 (10) and $\bar{G}Ct$ values of 10-100 (3). The wide ranges of these parameters may reflect genuine differences between waters (or wastewaters) but may also reflect different design approaches. Hudson (7) has suggested use of $\bar{G}t$ values of 2×10^5 which he claimed would produce high density floc with settling velocities equivalent to those of larger lower density floc produced at low $\bar{G}$ values. Camp (1) has suggested use of higher $\bar{G}$ values and resulting lower floc volumes to get equivalent primary particle agglomeration but with lower solids loadings on subsequent separation units.

Values in the ranges above are certainly ample for wastewater flocculation in flow through units. Because of the larger coagulant doses commonly used in wastewater treatment (especially with phosphorus removal) detention times and $\bar{G}t$ values can generally be lower. Culp et al (12) recommend a maximum of 15 minutes detention for wastewater coagulation. Culp and Culp (8) recommend using paddle speeds of 0.5 to 1.0 fpm.

Tapered flocculation in which the flow is exposed to decreasing $\overline{G}$ values as it passes through the unit, can provide a rapid build-up of small dense floc with subsequent agglomeration at lower $\overline{G}$ into larger but still dense particles. (9) (10) (11). Use of multi-compartment flocculators not only permits tapered flocculation, but also greatly reduces the high short-circuiting associated with single-compartment units. A wide variety of physical layouts are possible to achieve series flow through multiple compartments (10).

Flocculation units should have multiple compartments and should be equipped with adjustable speed mechanical stirring devices to permit meeting changed conditions. In spite of simplicity and low maintenance, non-mechanical, baffled basins are undesirable because of inflexibility, high head losses and large space requirements.

Mechanical flocculators may consist of rotary, horizontal-shaft reel units as shown in Fig. 6-2, rotary vertical shaft turbine units as shown in Fig. 6-3 or other rotary or reciprocating equipment. Features of these various type units are discussed and compared elsewhere (9) (10) (11).

Tapered flocculation may be obtained by varying reel or paddle size on horizontal common shaft units or by varying speed on units with separate shafts and drives. A typical series of $\overline{G}$ values for successive compartments would be 100, 50 and 20 fps ft. In most cases, equipment should provide overall Gt values up to 2 x 10^5 at maximum drive speed. Speed variation over a range of 1:3 or 1:4 should be possible (10).

$\overline{G}$ values are determined from the hp actually transmitted to the fluid (water hp). This should be distinguished from the total input hp which includes losses in motors, drives, bearings, etc. It should be noted that $\overline{G}$ is a mean value for the entire flocculator volume. Practical limits are set to localized high values at the flocculator blades or paddles by specifying peripheral speeds below about 2 fps.

In applications other than coagulation with alum or iron salts, flocculation parameters may be quite different. Lime precipitates are granular and benefit little from prolonged flocculation or very low terminal $\overline{G}$ values. At Lake Tahoe a detention of 4.5 min. proved adequate. Culp and Culp (8) recommend a minimum of 5 min. but as much as 10 min may be needed to assure complete dissolution and reaction of CaO. As in water softening practice, $\overline{G}$ values of 100 or more are desirable.

Polymers which already have a long chain structure may provide a good floc at low $\overline{G}$t values. Often the turbulence and detention in the clarifier inlet distribution is adequate.

Settling and effluent clarity in the activated sludge process can frequently be improved by controlled flocculation between the aeration tank and clarifier. Parker, et al. (14) showed that flocculation at $\overline{G}$ = 40 to 60 fps ft and detention of 20 to 30 min. could reduce the SS in aerator effluent (after settling) by some 45 to 55 percent. The benefits of flocculation depend on the level of turbulence in the aerator, and on the sludge age which affects the natural flocculating characteristics of the sludge. In the above study sludge with a sludge age of 10 days was better destabilized and benefited more from flocculation than did sludge with a

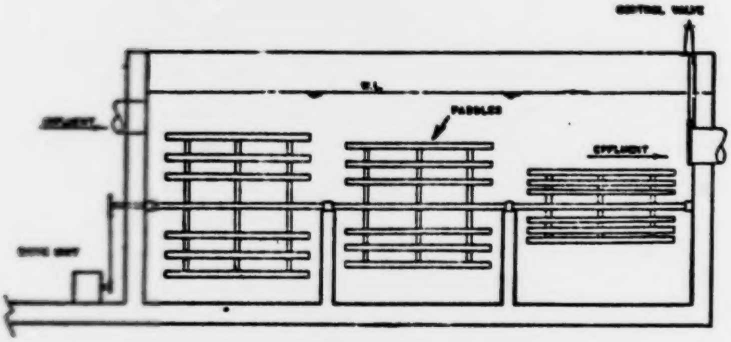
CONTROL VALVE
W.L.
PADDLES
INFLUENT
EFFLUENT
DRIVE UNIT

FIGURE 6-2
MECHANICAL FLOCCULATION BASIN
HORIZONTAL SHAFT-REEL TYPE

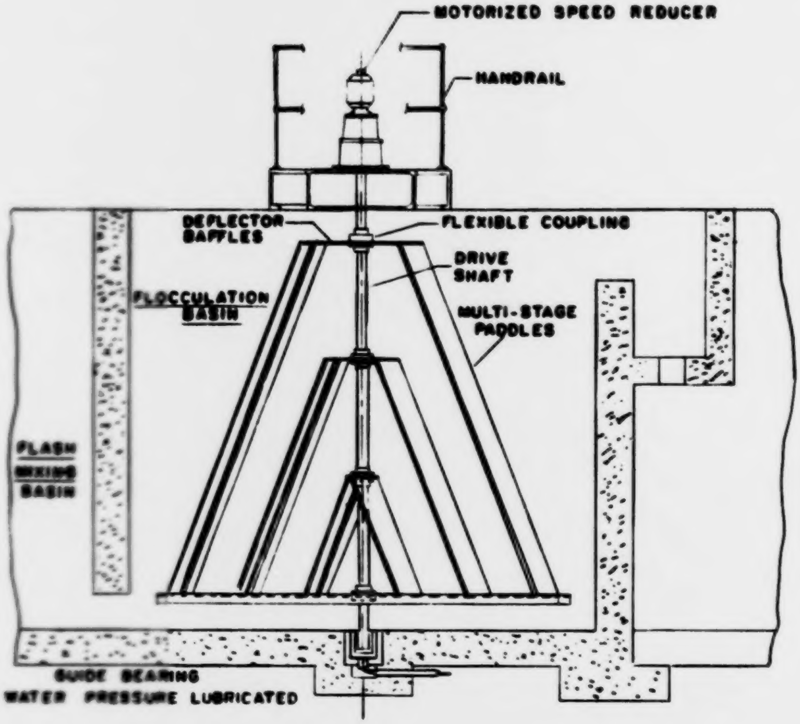
MOTORIZED SPEED REDUCER
HANDRAIL
DEFLECTOR BAFFLES
FLEXIBLE COUPLING
DRIVE SHAFT
FLOCCULATION BASIN
MULTI-STAGE PADDLES
FLASH MIXING BASIN
GUIDE BEARING
WATER PRESSURE LUBRICATED

FIGURE 6-3
MECHANICAL FLOCCULATOR
VERTICAL SHAFT PADDLE TYPE
(Courtesy of Ecodyne Corp.)

sludge age of 3 to 4 days or 12 days.

This behavior may be interpreted in light of the observation by Dean (15) that activated sludge contains an excess of natural anionic polymers. As sludge age increases these polymers are reduced first to levels where destabilization is better but then to levels below the optimum.

6.4 Solids-Contact Processes

Solids-contact processes combine chemical mixing, flocculation and clarification in a single unit designed so that a large volume of previously-formed floc is retained in the system. The floc volume may be as much as 100 times that in a "flow-through" system. This greatly increases the rate of agglomeration from particle contacts (11), and may also speed up chemical destabilization reactions.

Solids contact units are of two general types: slurry-recirculation and sludge-blanket. In the former, the high floc volume concentration is maintained by recirculation from the clarification to the flocculation zone, as illustrated in Fig. 6-4. In the latter, the floc solids are maintained in a fluidized blanket through which the wastewater under treatment flows upward after leaving the mechanically stirred-flocculating compartment, as depicted in Fig. 6-5. Some slurry-recirculation units can also be operated with a sludge blanket.

Solids-contact units have become popular in water treatment and are being increasingly considered in advanced wastewater treatment because of the following advantages:

1. Reduced size and lower cost result because flocculation proceeds rapidly at high floc volume concentration.

2. Single-compartment flocculation is practical because high reaction rates and the slurry effects overcome short circuiting.

3. Units are available as compact single packages, eliminating separate units.

4. Even distribution of inlet flow and the vertical flow pattern in the clarifier improve clarifier performance (16).

Equipment typically consists of concentric circular compartments for mixing, flocculation and settling. Velocity gradients ($\overline{G}$) in the mixing and flocculation compartments are developed by turbine pumping within the unit and by velocity dissipation at baffles. For ideal flexibility it is desirable to independently control intensity of mixing ($\overline{G}$) and sludge scraper drive speed in the different compartments. Ives (3) indicates that slurry-recirculation solids contact reactors in the water treatment field operate with a velocity gradient in the range of 60 to 120 fps ft. Hudson and Wolfner (11) indicate that in water treatment solid-contact reactors with variable-speed turbine-type agitators apply velocity gradients of 300 fps ft in the mixing zone while reaction zone values may vary from 100 fps ft near entrance to 20 fps ft at the settling zone boundary. Comparably proportioned units are being used in

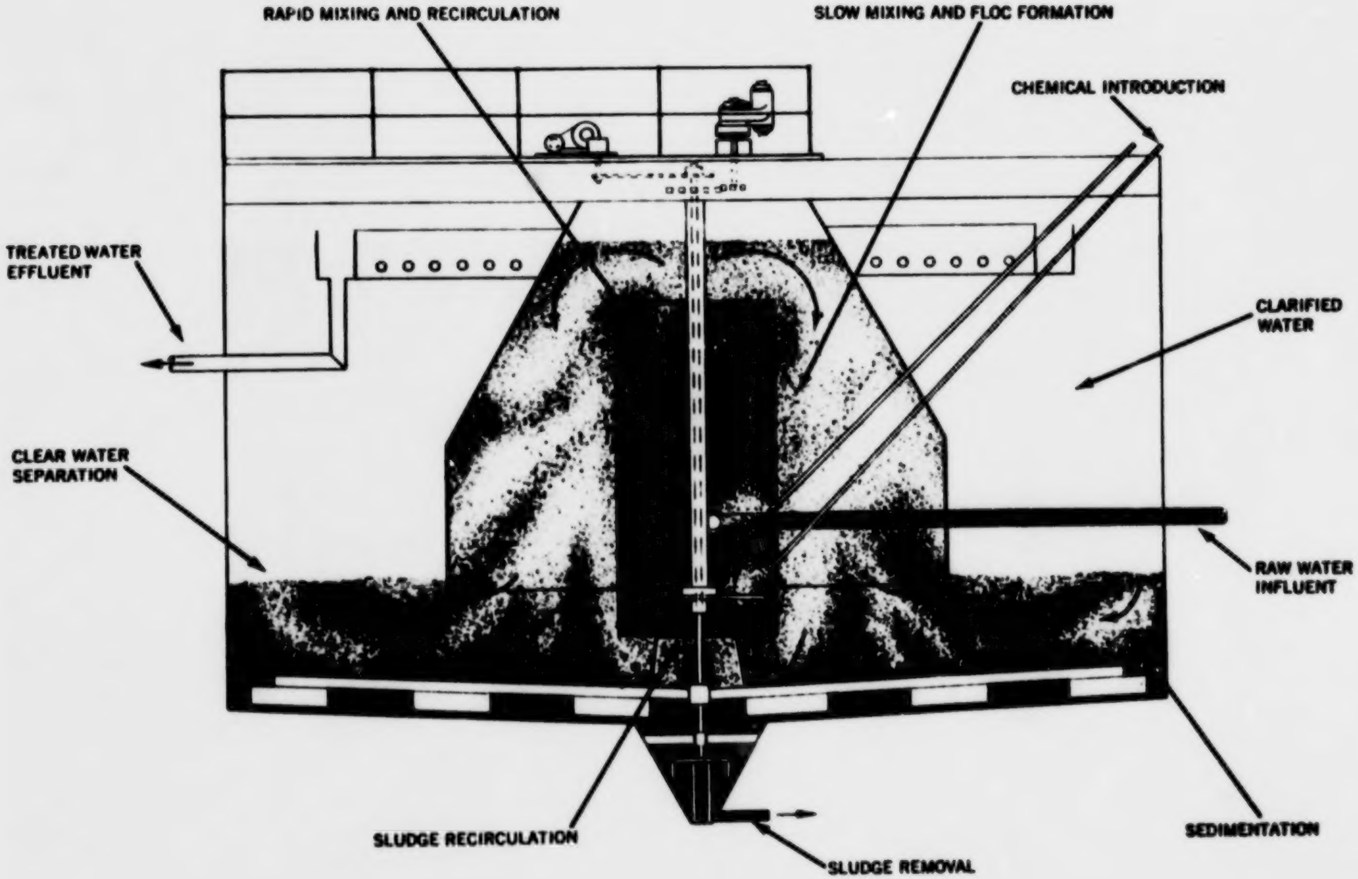
RAPID MIXING AND RECIRCULATION
SLOW MIXING AND FLOC FORMATION
CHEMICAL INTRODUCTION
TREATED WATER EFFLUENT
CLARIFIED WATER
CLEAR WATER SEPARATION
RAW WATER INFLUENT
SLUDGE RECIRCULATION
SEDIMENTATION
SLUDGE REMOVAL

FIGURE 6-4
SOLIDS CONTACT CLARIFIER WITHOUT SLUDGE BLANKET FILTRATION
(Courtesy of Ecodyne Corp.)

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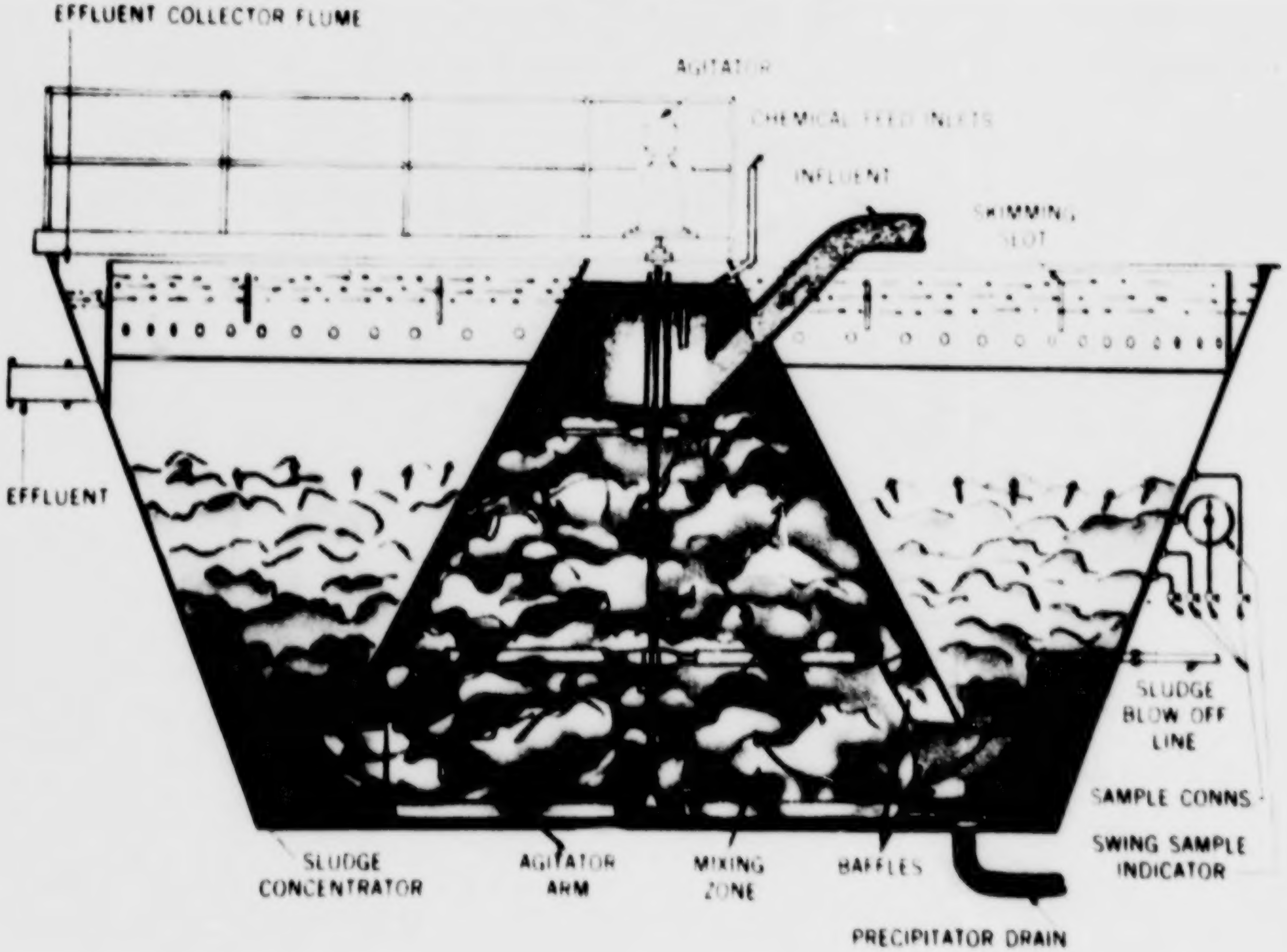
EFFLUENT COLLECTOR FLUME
AGITATOR
CHEMICAL FEED INLETS
INFLUENT
SKIMMING SLOT
EFFLUENT
SLUDGE BLOW OFF LINE
SAMPLE CONNS
SWING SAMPLE INDICATOR
SLUDGE CONCENTRATOR
AGITATOR ARM
MIXING ZONE
BAFFLES
PRECIPITATOR DRAIN

FIGURE 6-5
SOLIDS CONTACT CLARIFIER WITH SLUDGE BLANKET FILTRATION
(Courtesy of the Permutit Co.)

wastewater treatment, but with little explicit consideration of $\overline{G}$ values.

Experience with solids-contact units in wastewater treatment has up to now been limited to slurry-recirculation units. Culp and Culp (8) have expressed concerns about the use of sludge-blanket units: septicity and uncontrolled blanket upsets under varying-load conditions. Slurry-recirculation units not requiring sludge blankets or with minimum blanket depths are not very sensitive to such upsets. Units equipped with scrapers have operated without septicity problems treating secondary effluent at Nassau County, N.Y. (17) and at Ely, Min. (18).

Operation of slurry-recirculation solids-contact units is typically controlled by maintaining steady levels of solids in the reaction zone. For lime treatment of wastewater at Ely and Blue Plains a solids concentration of 10 to 12 percent by weight was found most effective (for phosphorus removal) (18) (19). For tertiary alum treatment at Nassau County 45 percent floc volume concentration proved most satisfactory (17).

Design features of solids-contact clarifiers should include:

1. Rapid and complete mixing of chemicals, feedwater and slurry solids must be provided.This should be comparable to conventional flash mixing capability and should provide for variable control of $\overline{G}t$ values, usually by adjustment of recirculator speed.

2. Mechanical means for controlled circulation of the solids slurry must be provided with at least a 3:1 range of speeds. The maximum peripheral speed of mixer blades should not exceed 6 ft/sec. Rushton and Mahoney (20) offer means of estimating pumping capacity of mixers.

3. Means should be provided for measuring and varying the slurry concentration in the contacting zone up to 50 percent by volume.

4. Sludge discharge systems should allow for easy automation and variation of volumes discharged. Mechanical scraper tip speed should be less than 1 fpm with speed variation of 3:1.

5. Sludge-blanket levels must be kept a minimum of 5 feet below the water surface.

6. Effluent launders should be spaced so as to minimize the horizontal movement of clarified water.

Most of the above requirements are based on those cited in Water Treatment Plant Design (10). Further considerations include skimmers and weir overflow rates. Skimmers should be provided on all units since even secondary effluents contain some floatable solids and grease. Overflow rates and sludge scraper design should conform to the requirements of other clarification units.

6.5 References

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CHAPTER 7

GRAVITY SEPARATION

7.1 Introduction

Gravity separation refers to the removal of SS whose specific gravity difference from that of water causes them to settle or rise during passage through a tank or basin under quiescent conditions. Separation by settling is termed sedimentation; separation by rising is termed flotation. The size of particles determines the fluid drag retarding this separation. For a given specific gravity, smaller particles having greater surface area encounter more drag and hence are more difficult to separate.

The factors affecting separation efficiency are discussed in depth for sedimentation, and separate sections cover each of its major applications. The section on flotation indicates special considerations pertaining to this process and deals with its applications. Finally two sections deal with devices which enhance the performance of sedimentation basins.

7.2 Configuration of Sedimentation Units

The tanks or basins in which sedimentation is carried out (also frequently termed clarifiers) may be classified as horizontal flow or vertical-flow according to the predominant direction of the flow path from inlet to outlet. It should be noted that, depending on placement of inlets and outlets, certain designs particularly small radial flow tanks may have a flow path with significant components in both horizontal and vertical directions.

7.2.1 Vertical-Flow Units

Vertical-flow applications in the U.S. have generally been limited to settling compartments in flocculation-clarifiers, solids-contact units and activated sludge systems of similar configuration (Aero-Accelator, Rapid Block, etc.). In Europe, vertical-flow basins have been used extensively. Kalbskopf has illustrated several European designs (1).

Vertical-flow units may be annular or rectangular, and are generally narrower at the bottom than at the top. In annular designs, the flow is distributed at the bottom along the circumference of the tank and rises to peripheral or radial effluent weirs or launders. Flow in rectangular tanks is distributed at the bottom along the length of the tank and rises to longitudinal or transverse effluent weirs or launders.

Annular units have been built with outside diameters to 150 ft, but the width from inner wall to outer wall is much less. Figures 6-4 and 6-5 illustrate annular, vertical-flow units.

7.2.2 Horizontal-Flow Units

In the U.S. horizontal-flow units, both rectangular and circular, are most often used for sedimentation applications. Tank proportions, inlet and outlet arrangements and types of sludge and scum collecting equipment are summarized and discussed in the ASCE/WPCF Manual for Sewage Treatment Plant Design (2). Individual bays of rectangular tanks should have a length to width ratio of at least four.

Flow through rectangular tanks enters at one end, passes a baffle arrangement, and traverses the length of the tank to effluent weirs. In narrow tanks, longitudinal collectors scrape sludge to single or multiple hoppers at one end (Figure 7-1). In tanks with multiple wide bays, the longitudinal collectors scrape sludge to a cross collector which then moves the sludge to a central hopper. Circular designs employ three inlet/outlet configurations with corresponding flow paths as shown in Figure 7-2. In configurations 7-2 (a) and (c), sludge is removed by mechanical scraping to a central hopper or draw-off. In configuration 7-2 (b) a hydraulic suction sludge removal system is employed.

7.3 Basic Factors Affecting Settling Tank Design

7.3.1 Hydraulic Loading

The basic parameter to which settling tank performance is related is the surface hydraulic loading (Q/A). This is the inflow (Q) divided by the surface area (A) of the basin, and is commonly expressed in units of gpd/sq ft.

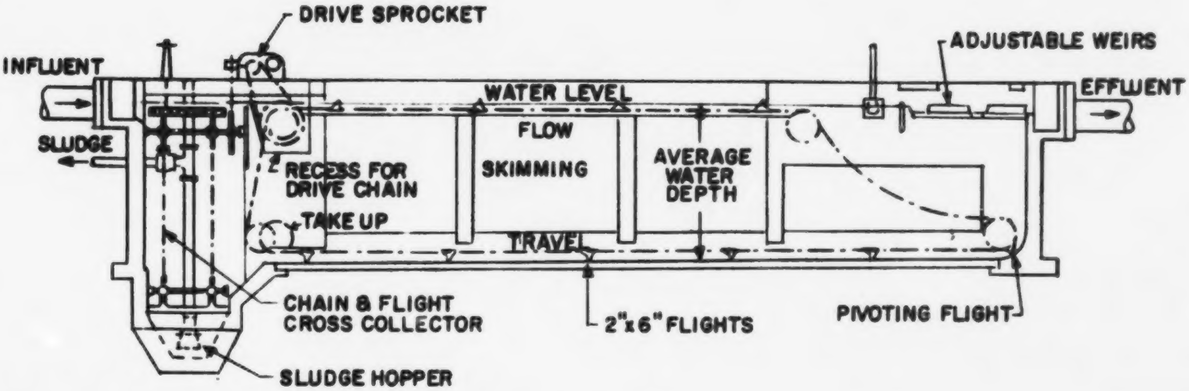
Hazen (3) showed that under the following assumptions performance is a function of surface loading alone:

1. Quiescent or non-turbulent flow
2. Uniform distribution of velocity over all sections normal to general flow direction
3. Discrete non-interacting particles
4. No resuspension of particles once they reach the floor of the basin

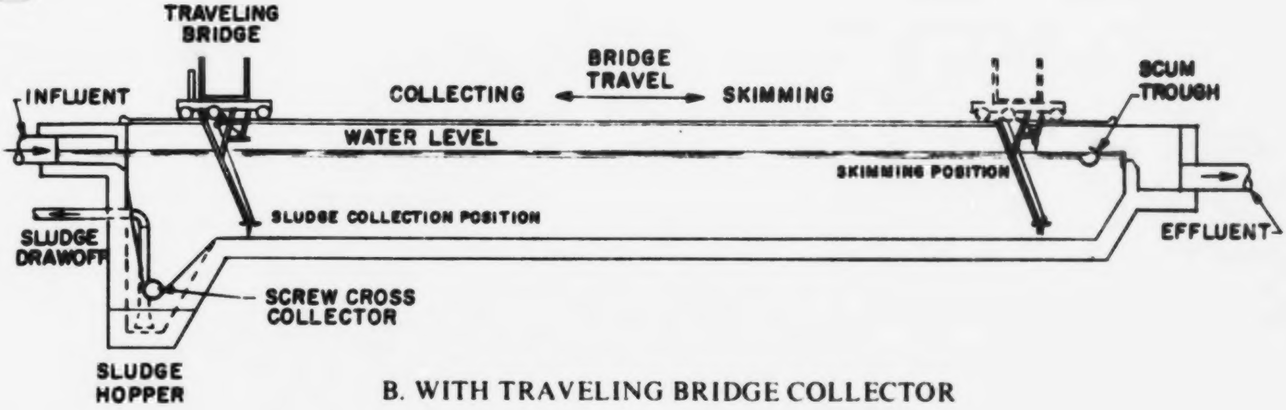
Under these conditions all particles whose settling velocity (Vs) exceeds Q/A are removed. In addition, in horizontal flow tanks particles of lower settling velocities are partially removed in the proportion Vs/(Q/A).

In actual basins conditions depart in many respects from those assumed in Hazen's original analysis. The most significant of these departures are:

1. Currents induced by inlets, outlets, wind and density differences may cause short circuiting or dead spaces within the tank.
2. Turbulence due to forward velocity or currents in the tank retards settling.
3. Flocculent solids may agglomerate into larger particles during passage through the basin.


DRIVE SPROCKET
ADJUSTABLE WEIRS
INFLUENT
EFFLUENT
WATER LEVEL
FLOW
SLUDGE
RECESS FOR DRIVE CHAIN
SKIMMING
AVERAGE WATER DEPTH
TAKE UP
TRAVEL
CHAIN & FLIGHT CROSS COLLECTOR
2"x6" FLIGHTS
PIVOTING FLIGHT
SLUDGE HOPPER

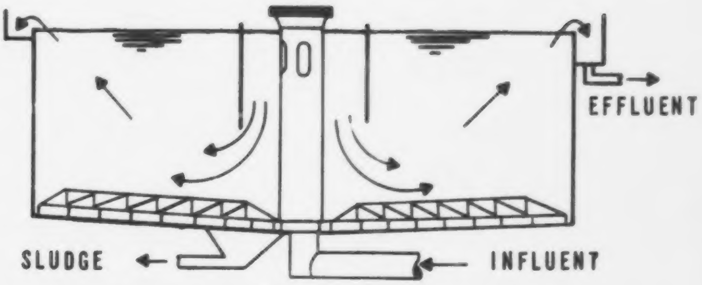
A. WITH CHAIN AND FLIGHT COLLECTOR


TRAVELING BRIDGE
BRIDGE TRAVEL
COLLECTING
SKIMMING
SCUM TROUGH
INFLUENT
WATER LEVEL
SKIMMING POSITION
SLUDGE COLLECTION POSITION
SLUDGE DRAWOFF
EFFLUENT
SCREW CROSS COLLECTOR
SLUDGE HOPPER

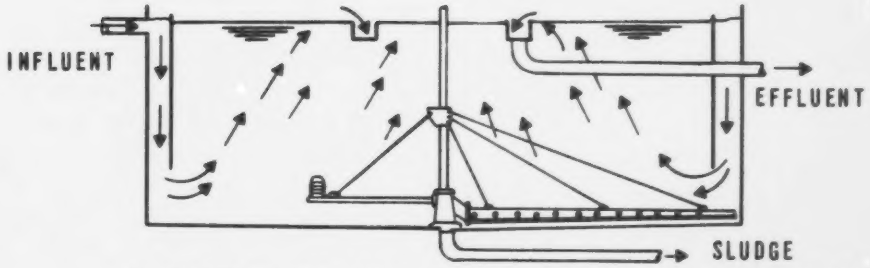
B. WITH TRAVELING BRIDGE COLLECTOR

FIGURE 7-1

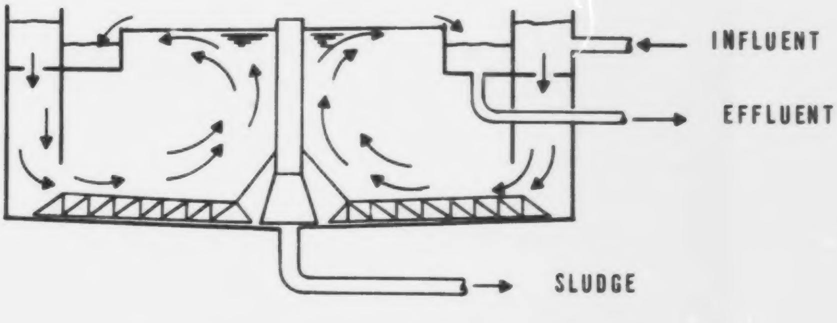
RECTANGULAR SEDIMENTATION TANKS
(Courtesy of FMC Corp.)


EFFLUENT
SLUDGE
INFLUENT

(a) CIRCULAR CENTER-FEED CLARIFIER WITH A SCRAPER SLUDGE REMOVAL SYSTEM


INFLUENT
EFFLUENT
SLUDGE

(b) CIRCULAR RIM-FEED, CENTER TAKE-OFF CLARIFIER WITH A HYDRAULIC SUCTION SLUDGE REMOVAL SYSTEM


INFLUENT
EFFLUENT
SLUDGE

(c) CIRCULAR RIM-FEED, RIM TAKE-OFF CLARIFIER

FIGURE 7-2

TYPICAL CLARIFIER CONFIGURATIONS

4. Sludge may be scoured and resuspended at high forward velocities.

5. When influent solids concentrations are high, particles settle as a mass rather than discretely.

The subsections below indicate how investigators, most notably Camp (4), have attempted to account for these departures by relating performance to additional parameters. The relationships are not generally adequate to permit prediction of performance from design values of the parameters, but they do provide insights helpful in deciding a number of tank features such as shape, depth, inlet type, etc. In addition, such relationships offer guidance in translating settling test results into sizing for full scale tanks. Procedures for conducting and interpreting such tests have been outlined by O'Connor and Eckenfelder (5) and others (6) (7) (8).

To account for departures of full scale tanks from ideal or test conditions scale-up factors in the following ranges have been suggested (5):

Sizing Parameter	Scale-Up Factor
Area	1.25 to 1.75
Volume	1.5 to 2.0

These scale-up factors are not intended to cover extreme variations in flows or solids loadings, or to allow for operation at temperatures significantly different from those in the tests. Neither do they include standby capacity as needed for units critical to overall plant performance. Smith (9) has discussed the use of excess capacity factors to provide for standby and to cover expected variations in loadings.

7.3.2 Short Circuiting

Short circuiting can greatly reduce the removal efficiency of a settling tank. Effects are most critical for flocculent suspensions whose removal is affected by detention time (Sec. 7.3.4), but depending on the current pattern, removal of discrete particles may also be affected. Short circuiting is accentuated by high inlet velocities, high outlet weir rates, close placement of inlets and outlets, exposure of tank surface to strong winds, uneven heating of tank contents by sunlight, and density differences between inflow and tank contents. Density-induced short circuiting can be a significant factor in secondary settling tanks handling activated sludge mixed liquor (10). Inlet and outlet conditions, tank geometry, and density differences due to influent SS concentrations produce steady short circuiting, whereas effects of other factors are generally intermittent and unpredictable.

The degree of short circuiting can be measured using tracer studies. Figure 7-3 shows results of such studies on four types of settling tanks (11), where short circuiting was due primarily to inlet and outlet conditions and tank geometry. Studies of this type have confirmed that such short circuiting is minimized in narrow, rectangular, horizontal-flow tanks and is most serious in circular horizontal flow tanks. Although upflow tanks show the least short

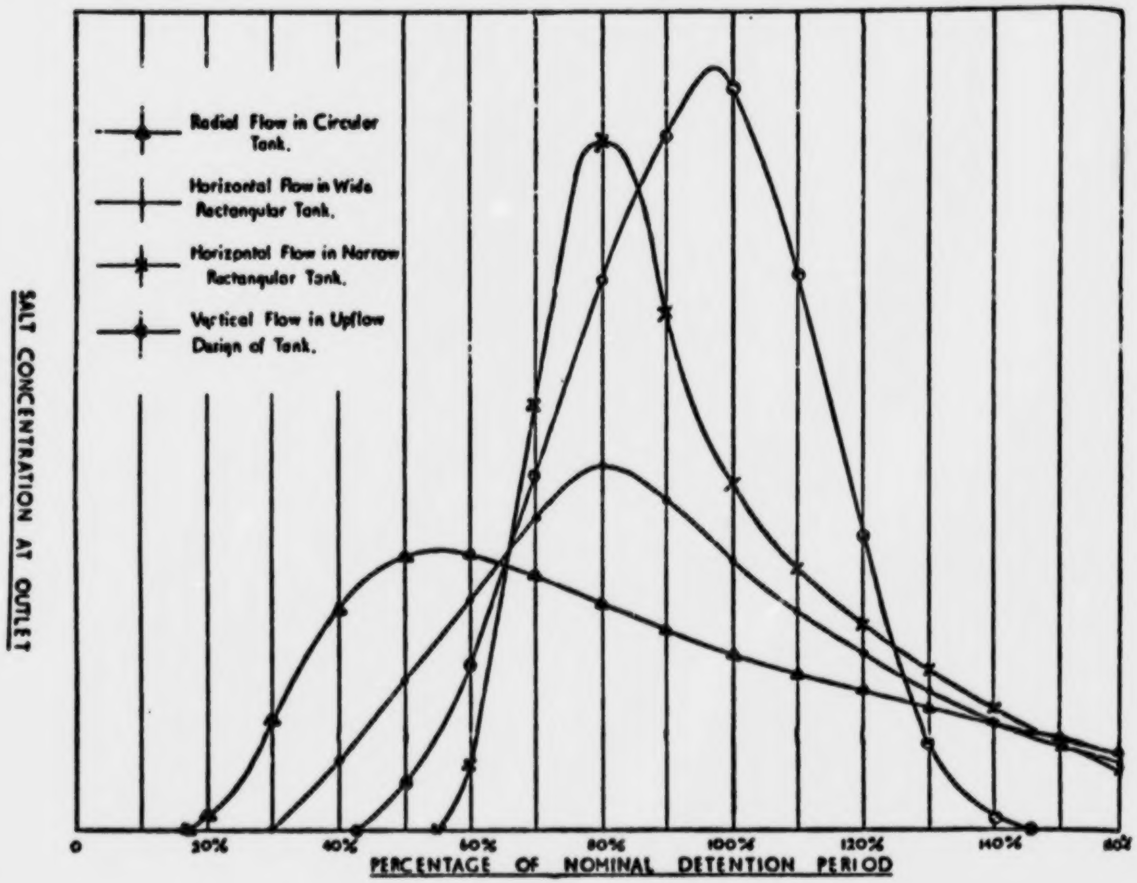


FIGURE 7-3

RESULTS OF SALT-INJECTION TESTS WITH DIFFERENT TYPES OF SEDIMENTATION TANKS circuiting, practical problems in obtaining uniform initial flow distribution have limited their use to small diameter units.

The degree of short circuiting in circular units can vary considerably, however, depending on the type of inlet used. Inlet conditions have been shown to be more critical than those at the outlets (12). For activated sludge final settling tanks, peripheral feed and certain special-design center feed inlets have been shown to cause less short circuiting than conventional center feed inlets (10) (13) (14).

Even where the degree of short circuiting can be measured or predicted, techniques for evaluating the effect on tank performance (1) (15) are questionable as to their utility and accuracy. Hence, the best design approach is to avoid short circuiting as far as possible, thus minimizing uncertainty as to its effects. The most important factors to consider in controlling short circuiting are dissipation of inlet velocity, protection of tanks from wind sweep and uneven heating, and reduction of density currents associated with high inlet SS concentrations (13).

Such density current short curcuiting is a particular problem in settling tanks for activated sludge. Fitch (10) has presented estimates of the velocities of such currents as a function of SS concentration, and has compared two fundamental approaches to preventing short circuiting from this source. These are dynamic stabilization as proposed by Camp (4) and density stabilization. Dynamic stabilization requires shallow basins with high forward velocities. (Froude numbers of 5×10^{5} or greater). The resulting friction losses, in theory, counteract stratification and instability of flow. Density stabilization essentially establishes an upflow type pattern by introducing the dense feed at low velocities and close to the tank bottom. Fitch showed that low inlet velocities are essential to successful density stabilization, and proposed a novel center inlet design to achieve such velocities (see Section 7.4).

7.3.3 Turbulence

Turbulence levels in a settling basin are normally difficult to estimate. The only exception is turbulence due to drag from net forward velocity. Camp (4) has presented a basis for estimating turbulence from this source and for compensating for its effects by increasing tank area. Required increases vary directly with forward velocity in the tank and with the desired removal rate.

Good design practice is to minimize other sources of turbulence such as inlet, outlet, wind and density currents. These sources produce unpredictable levels of turbulence and may increase short circuiting. Even where the degree of turbulence during sedimentation can be definitely measured the effect on removal of flocculent particles is not easily predicted, because agglomeration induced by turbulence can alter particle sizes and localized settling velocities.

7.3.4 Particle Agglomeration

For the flocculent suspensions handled in wastewater treatment, particle contact and agglomeration continues during sedimentation. Two mechanisms produce particle contacts: velocity gradients within the settling tank, and differential settling rates; each of which permit faster moving particles to overtake slower ones. Depending on the nature of the influent suspension, either mechanism can significantly affect both the size and settling velocity of floc and the fraction of fine, unsettleable particles remaining in suspension. Regardless of surface loading on a settling tank, attachment of smaller, unsettleable particles onto larger ones of separable size is essential in attaining high SS removal efficiences. In any case, these larger particles must have the opportunity to agglomerate to sizes which will be removed at the maximum surface loadings applied to the tank. Otherwise massive failure of the separation process will occur with significant loss of SS in the effluent.

Camp (4) asserted that the rate of particle contacts due to differential settling depends only on the characteristics of the suspension. Fitch (16) (17) maintained that the rate also increased with tank depth. In either case, the total number of contacts occurring due to differential settling is a direct function of detention time, which at a given surface hydraulic loading is, in turn, a function of settling tank depth. In contrast, the rate of particle contacts due to velocity gradients increases with forward velocity and hence decreases with depth.

For the 10 to 15 ft tank depths normally used in wastewater treatment in the U.S., agglomeration depends mainly on differential settling. For wastewaters such as raw sewage, which agglomerate slowly under differential settling, detention time can have a significant effect on settling tank performance (See Section 7.5).

Camp (4) urged the use of much shallower settling tanks, theorizing that the higher velocity gradients would accelerate particle agglomeration sufficiently to more than offset the reduction in detention time. Fitch (16) disputed this noting that in stirred settling tests velocity gradients comparable to those proposed by Camp provided little flocculation. In any case, common U.S. practice has remained to design fairly deep tanks with low forward velocities (about 1 fpm at mean flow) and to depend on some other means than gradients due to forward velocity to achieve desired flocculation. Kalbskopf (1) indicated that in Europe it is common, to design shallower (3 to 10 ft depth) primary settling tanks with higher forward velocities (2.5 fpm at mean flow and up to 7.5 fpm at maximum flow). Studies for the Emscher Mouth treatment facility (18) showed only minor variation of primary effluent SS with forward velocity. Performance related much more to surface loading. For any given surface loading, however, the best performance was at a velocity in the range between 1.6 and 2.5 fpm.

It is well recognized that increasing velocity gradients by stirring the inlet zone of a settling tank can often improve performance (1) (4) (19). Essentially this combines mechanical flocculation and settling in a single tank. Compartmentation is desirable to reduce short circuiting. The major advantage of such combined units is that a suspension can be flocculated at decreasing $\overline{G}$ values (see Section 6.1) down to very low levels and then delivered to sedimentation without subjecting the suspension to the shearing effects of collection and redistribution. Recognizing this advantage, several equipment manufacturers offer combined units designed on this basis. Where flocculation is to be used to upgrade performance of existing settling tanks, the possibility of locating flocculation mechanisms directly in the tanks should be considered. (See U.S. EPA, Process Design Manual for Upgrading Existing Wastewater Treatment Plants).

7.3.5 Bottom Scour

Where high forward velocities are used, the possibility of scouring previously deposited sludge should be analyzed. As a rule of thumb, forward velocities should be limited to from 9 to 15 times the settling velocity of critical size solids to avoid scour (20).

7.3.6 Hindered Settling and Compaction

When a concentrated suspension such as activated sludge mixed liquor settles under quiescent conditions, a distinct interface develops almost immediately between the sludge and the clarified liquid above it (21). As illustrated in Figure 7-4, this interface subsides for a time at a constant rate. This rate is termed the initial settling velocity of the sludge. Because the accompanying upward displacement of liquid reduces this settling velocity to below that of discrete particles of the sludge, the process is termed hindered settling.

As the sludge mass continues to settle an interparticle structure develops in the more concentrated lower layers and the subsidence rate slows further. Figure 7-5 illustrates this compaction or thickening of sludge in a full scale tank. If high sludge concentrations are to be obtained, thickening rather than solids separation may control the tank sizing. Sizing secondary settling tanks for activated sludge to meet thickening requirements is discussed in Section 7.6.

7.4 Clarifier Design Considerations

7.4.1 General

In selecting the particular tank shape, proportions, equipment, etc. the designer should:

1. Provide for even inlet flow distribution in a manner which minimizes inlet velocities and short circuiting.

2. Minimize outlet currents and their effects by limiting weir loadings (see Sec. 7.5 and 7.6) and by proper weir placement.

3. Provide sufficient sludge storage depths to permit desired thickening of sludge.

4. Provide sufficient wall height to give a minimum of 18 inches of freeboard.

5. Reduce wind effects on open tanks by providing wind screens and by limiting fetch of wind on tank surface with baffles, weirs or launders.

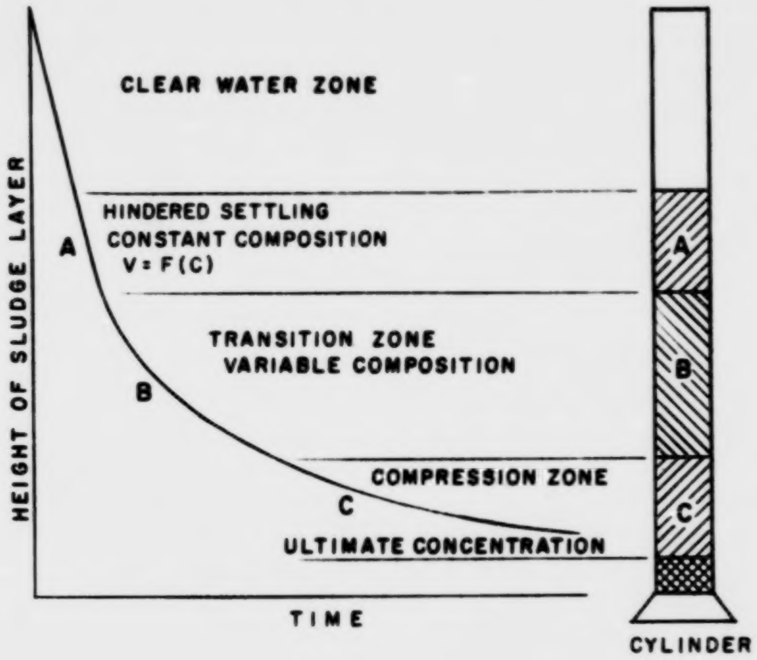

CLEAR WATER ZONE
HINDERED SETTLING
CONSTANT COMPOSITION
V = F(C)
A
TRANSITION ZONE
VARIABLE COMPOSITION
B
COMPRESSION ZONE
C
ULTIMATE CONCENTRATION
HEIGHT OF SLUDGE LAYER
TIME
CYLINDER

FIGURE 7-4

SCHEMATIC REPRESENTATION OF SETTLING ZONES

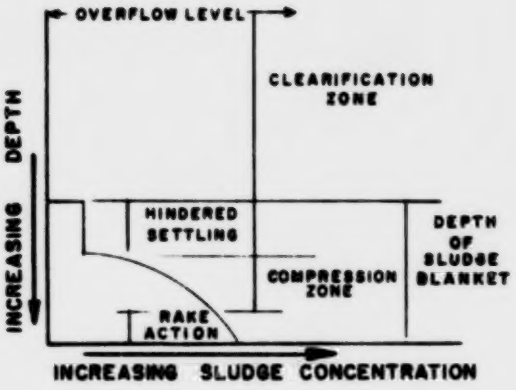

OVERFLOW LEVEL
CLEARIFICATION ZONE
INCREASING DEPTH
HINDERED SETTLING
COMPRESSION ZONE
DEPTH OF SLUDGE BLANKET
RAKE ACTION
INCREASING SLUDGE CONCENTRATION

FIGURE 7-5

SEDIMENTATION IN A SECONDARY SETTLING TANK

6. Consider economy of alternative layouts which can be expected to provide equivalent performance.

7. Maintain equal flow to parallel units. This is most important and often forgotten. Equal flow distribution between settling units is generally obtained by designing equal resistances into parallel inlet flow ports or by flow splitting in symmetrical weir chambers.

7.4.2 Inlet Design

Inlet design for rectangular tanks, where the distance from inlet to outlet is large, is less critical than for circular tanks where there is generally little separation between inlet and outlet.

In rectangular tanks flow is distributed over the width of the tank by provision of multiple inlets. Size and spacing vary considerably from one design to another. Small openings are avoided in wastewater applications because of the possibility of fouling. Maximum spacings are generally less than 10 ft. Target baffles are commonly provided to help dissipate the velocity of the inlet jets. Distribution to multiple inlets in a rectangular tank usually involves a manifold conduit. A method, developed by Dobbins, for design of inlets and manifold conduits, is presented elsewhere (22).

The common type of center feed for circular tanks depends on symmetrical baffling to distribute flow equally in all radial directions. The high degree of short circuiting with such inlets has led manufacturers to develop several special inlet designs for circular tanks both center and peripheral feed.

Figure 7-2b and c show peripheral feed units. In these units, inlet ports discharge outside a deep peripheral baffle and flow passes under this baffle to enter the tank. In a peripheral feed unit manufactured by Lakeside Equipment Corp., the inlet line to the tank discharges tangentially into a tapered race located behind a similar skirt baffle. The manufacturer claims that the tangential motion imparted to the tank contents reduces short circuiting. In model studies, the latter type of peripheral unit showed significantly higher removals of iron floc than a similarly loaded center feed unit (23). This was attributed to better conditions for particle agglomeration in the peripheral feed model.

A center feed inlet manufactured by Dorr-Oliver, Inc. has two races with tangentially-introduced flows rotating in opposite directions. Shear between these rotating flows dissipates the energy of the inlet velocity before the inflow leaves the feedwell (10).

The modular Energy Dissipating (MED) Feedwell, manufactured by Envirotech Corp., forces all flow to pass through honeycombs of small tubes, mounted vertically around the entire feedwell periphery. The manufacturer claims that the honeycomb creates a laminar flow pattern with uniform radial velocities and that periodically reversing the modules on their pivot mountings(changing flow direction through them) will clear the honeycomb of any accumulated solids.

Available data comparing performance of primary and secondary clarifiers using special and conventional inlets are presented in Table 7-1.

7.4.3 Economy

The two major elements in settling tank cost are the structure and the sludge collection mechanism. Installed cost for the mechanism is typically 30 to 40 percent of the structural cost. Structural costs for multiple rectangular and circular tanks (horizontal flow) are comparable, provided common-wall construction is used for the rectangular units (1) and liquid depths are not more than about 10 ft. At greater depths, circular units with tank walls designed as hoops show increasing savings. Single circular units are less expensive than the same size rectangular basin. Where tanks must be covered, costs may favor rectangular units because of their shorter roof spans. European data (1) indicate structural costs for vertical flow units may run 50 percent higher than horizontal units of the same volume, but the vertical tanks have deep conical bottoms eliminating the need for costly sludge collector mechanisms.

Rotary collectors for circular tanks generally cost 20 percent less than chain-and-flight collectors for comparable rectangular units. In addition, maintenance requirements for the rotary units are decidedly lower. Travelling-bridge collectors for rectangular tanks apparently compete favorably with rotary circular collectors in cost and ease of maintenance. They are common in Europe but until recently have not found widespread application in the U.S. A recent comparison (27) for secondary tanks showed a floating travelling bridge collector with siphon sludge drawoffs to be decidedly cheaper than either chain-and-flight or circular mechanisms.

7.4.4 Skimming

In rectangular tanks with chain-and-flight collectors, skimmings are moved toward their discharge point by return travel of the flights at the tank surface (Figure 7-1A). In circular tanks skimmings are moved by travel of a surface arm attached to the rotary collector (Figure 7-2). A surface arm can be similarly used in rectangular tanks with travelling-bridge collectors. Discharge of scum from the settling tank may be continuous or intermittent depending on quantity produced. Skimmer mechanisms are of two types: dipping-weir and sloping-beach. In the first, a slotted tilting pipe or other weir device is positioned during skimming so that scum overflows from the tank together with considerable water. In sloping-beach units scum is raked mechanically up a beach leaving most of the water behind. The latter are simple to provide on circular tank mechanisms where they are almost standard. For rectangular tanks with chain-and-flight collectors, however, a separate mechanism is required to move scum up the sloping beach. In this application, use of sloping beach rather than weir type it is desirable to minimize the moisture content in the scum and facilitate subsequent handling. Where scum is to be pumped away from the tanks the less expensive weir-type skimmers are generally preferable.

TABLE 7-1

PERFORMANCE OF SPECIAL SETTLING TANK INLETS

Special Inlet Type and Application	Location	Test Period	Loading		Effluent SS		Reference
			Special Inlet gpd/sq ft	Conventional Center Feed gpd/sq ft	Special Inlet mg/l	Conventional Center Feed mg/l	
Peripheral Feed (Rex-Nord) – Activated Sludge Final Clarifier	Ann Arbor, Mich.	5/22/61 to 6/22/61	1408	951	11	14	24
	Sioux Falls, S.D.	8/12/58 to 9/27/58	2015	401	30	30	24
Peripheral Feed (Lakeside Equipment) Primary Clarifier	Ewing-Lawrence, N.J.	6/70 to 12/70	1000±	1000±	67	81	25
Modular Cell Inlet Feedwell (Envirotech)	Odgen, Utah	10/70 to 5/71	520-715	520-715	31	46	26
			850-950	850-950	31	48	26
			950-1150	950-1150	31	54	26

7.5 Primary Sedimentation

In theory, sizing of primary tanks may be regarded as a question of economics. Successive increments of tank area (providing lower loadings and longer detentions) typically yield diminishing returns in performance. At some point it becomes more economical to accept higher loads in subsequent units rather than provide more primary tank capacity. Some designs have even omitted primary tanks entirely. Von der Emde (28) has indicated this may be advantageous if one or more of the following conditions apply:

1. Sludge from the facility is to be pumped away for treatment elsewhere
2. Problems are expected with odors in primary tanks or poorly settling sludge in secondary tanks
3. Aerobic digestion or extended aeration processes are to be used.

As a practical matter, performance-loading relationships adequate for use in cost optimization studies can currently be obtained only by extensive testing of the actual wastewater in existing full scale or pilot facilities, taking into account variations in flows and characteristics. The generalized performance-loading curves for sedimentation units available in the literature (2) (8) (19) (29) are unsatisfactory even as a basis for predicting performance at particular design loadings much less for cost optimization studies. Such curves are based on average daily plant flows from diverse sources, and ignore effects of diurnal flow variations and of major in-plant flows such as waste secondary sludge, which may be recycled to the primary tanks. The effect of such unaccounted-for factors may be seen in the wide scatter of removal-loading data plotted in the WPCF ASCE Sewage Treatment Plant Design Manual (2).

In the absence of reliable performance-loading relations, primary tank designs may be based on the typical parameters shown in Table 7-2.

TABLE 7-2

TYPICAL DESIGN PARAMETERS FOR PRIMARY CLARIFIERS

Type of Treatment	Hydraulic Loading Average	 Peak	Depth
	gpd sq ft		ft
Primary Settling Followed by Secondary Treatment	800-1,200	2,000-3,000	10-12
Primary Settling with Waste Activated Sludge Return	600-800	1,200-1,500	12-15

Sizing should be calculated for both average and peak conditions (if flow equalization is not used) and larger size used.

These parameters are applicable to normal municipal wastewaters primarily of domestic origin and should provide SS removals of 50 to 60 percent.

Weir loading limitations between 10,000 and 30,000 gpd ft (24-hr basis) have been suggested for primary tanks (19) (29). At usual surface loadings, up to 1200 gpd sq ft, round tanks with single peripheral weir fall in this range for all but very large diameters (> 100 ft). Thus normal practice is to provide only the single weir. In contrast, at surface loadings as low as 600 gpd sq ft rectangular tanks with single transverse weirs across the effluent end exceed this range if the tank length is over 50 ft. Although rectangular tanks with weir rates of more than 100,000 gpd ft have shown SS removal in the normal range (30), rectangular tanks are commonly equipped with multiple weir troughs to provide loadings of 30,000 gpd ft or less. However, weir loadings are not as critical for primary tanks as they are for secondary clarifiers.

Sludge solids can be estimated directly from the expected SS removal, making sure to include waste activated sludge returned to the primary tank in the solids loading. The sludge volume can be calculated based on expected concentration. If sludge is properly thickened in the primary tank and pumping is carefully controlled to avoid pulling excess water, solids concentrations of 2 to 7 percent may be obtained. On this basis typical primary sludge volumes for domestic sewage would range from 0.2 to 0.5 percent of plant flow. The concentration used in particular estimates should be based on actual plant experience or at least on settling thickening tests. Quantities of skimmings are quite variable. On a sustained basis few plants average over 1 cu ft mg (free water decanted) but scum handling facilities should be capable of moving peak loads of perhaps six times this amount.

7.6 Secondary Sedimentation

7.6.1 Tank Sizing General

The approach to sizing secondary clarifiers varies with the type of biological process they serve.

7.6.1.1 Tank Sizing For Trickling Filter Effluent

Clarifiers following trickling filters are basically sized on hydraulic loading. Solids loading limits are not involved in this sizing. Where further treatment follows sedimentation, cost optimization may be considered in sizing the settling tanks, but the effort of developing adequate performance loading relations is seldom justified. Typical design parameters for clarifiers following trickling filters are presented in Table 7-3. In applying the hydraulic loading values from the table to design, sizing should be calculated for both peak and average conditions and the largest value determined should be used. At the indicated hydraulic loadings, settled effluent quality is limited primarily by the performance of the biological reactor not of the settling tanks.

TABLE 7-3

TYPICAL DESIGN PARAMETERS FOR SECONDARY CLARIFIERS

Type of Treatment	Hydraulic Loading		Solids Loading*		Depth
	Average	Peak	Average	Peak	
	gpd sq ft		lb solids day sq ft		ft
Settling Following Trickling Filtration	400-600	1,000-1,200	—	—	10-12
Settling Following Air Activated Sludge (Excluding Extended Aeration)	400-800	1,000-1,200	20-30	50	12-15
Settling Following Extended Aeration	200-400	800	20-30	50	12-15
Settling Following Oxygen Activated Sludge with Primary Settling	400-800	1,000-1,200	25-35	50	12-15

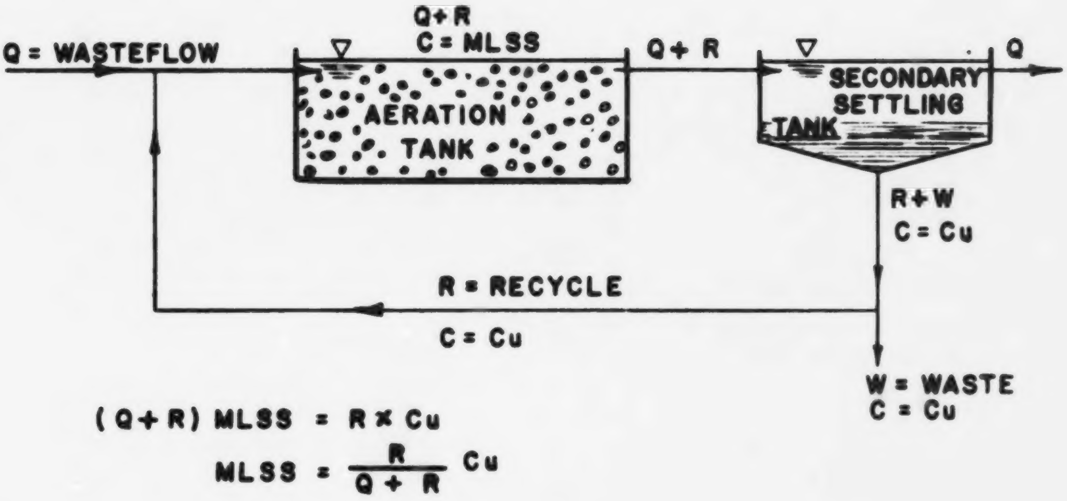
*Allowable solids loadings are generally governed by sludge thickening characteristics associated with cold weather operations.

7.6.1.2 Tank Sizing For Activated Sludge Mixed Liquor

Activated sludge settling tanks have two distinct functions: solids separation and production of a concentrated return flow to sustain biological treatment. Figure 7-6 illustrates how important final tank underflow concentration is in maintaining the level of active solids (and hence treatment) in the aerator mixed liquor.

As indicated in Section 7.3.6, the initial separation of activated sludge solids involves hindered rather than discrete settling. For this type of settling, tanks must be sized so the maximum surface hydraulic loading is less than the minimum initial settling velocity (ISV) expected at maximum mixed liquor concentration and at minimum temperature. If the hydraulic loading exceeds the ISV massive failure and overflow of solids will result.

To perform properly while producing a concentrated return flow, activated sludge settling tanks must be designed to meet thickening as well as solids separation requirements. The critical element in thickening is the rate at which solids are transported downward and removed in the tank underflow. This is termed the solids transport or solids flux capacity, generally expressed in the units of solids loading, lb sq ft day. When the actual solids loading applied to a tank exceeds its transport capacity, solids are being added faster than


Q+R
C = MLSS
Q = WASTEFLOW
AERATION TANK
Q+ R
Q
SECONDARY SETTLING TANK
R+W
C = Cu
R = RECYCLE
C = Cu
W = WASTE
C = Cu

$$(Q+R)\ MLSS = R \times Cu$$

$$MLSS = \frac{R}{Q+R}\ Cu$$

FIGURE 7-6

DEPENDENCE OF MLSS CONCENTRATION ON SECONDARY SETTLING TANK UNDERFLOW CONCENTRATION (31) they are being removed. If this condition persists the blanket of solids in the tank will build up and eventually overflow with drastic effects on effluent quality. If significant solids are lost from the system, biological treatment efficiency will be impaired. Tank depth may be important in containing blanket buildup from diurnal peaks in solids loading.

Dick (31) has analyzed solids transport capacity assuming both solids and (vertical) underflow velocity uniformly distributed over the plan area of the tank. Although these conditions are approximated only in moderate size circular tanks, this analysis provides major insight into the thickening process and represents the only rational and straightforward approach currently available for estimating solids transport capacity. Under the conditions assumed, solids transport capacity depends on only two factors: the thickening characteristics of the solids (i.e., the relation between subsidence rate and concentration within the sludge blanket) and the tank underflow rate. To get a concentrated underflow requires a low sludge return rate which in turn means low solids transport capacity. High underflow rates have been resorted to for handling poorly compacting sludges. This is only partly effective since while increasing solids transport capacity, higher underflow rates also increase solids loadings due to the higher sludge recycle.

Methods for developing hydraulic and solids loading parameters from tests of settling and thickening characteristics are discussed in Section 7.6.3. Typical design parameters for clarifiers in activated sludge systems treating domestic waste are given in Table 7-3. In applying hydraulic and solids loading values from this table, sizing should be calculated for both peak and average conditions and the largest value determined should be used.

Settling tests provide worthwhile guidance in selecting design loadings. They should certainly be included wherever pilot study of biological treatment is warranted by unusual waste characteristics or treatment requirements. Testing is essential in any case where proposed loadings go beyond the upper limits shown in Table 7-3.

Sizing activated sludge settling tanks according to proper hydraulic and solids loading parameters protects against massive failure, but does not by itself guarantee high quality effluent. After separation of the mass of activated sludge solids, significant quantities of small, slowly settling particles may still be left in the clarified liquor. The amount and character of such residual suspended solids logically relate to the loading and operating conditions in the aeration tank, but few specific studies have explored such relations. A study in Baltimore, Md. (32) indicated that sludges with poorer thickening characteristics left lower residual solids in the effluent. This was confirmed by studies covering a number of plants in Sweden (33).

As noted in Chapter 6, the concentration of these residual solids can be reduced by flocculation of the mixed liquor between aeration and settling, or by use of recirculation-type or sludge-blanket-type solids contact reactors. Finally, although rates are lower, flocculation in the clear water zone still appears to be a significant mechanism in removal of solids not already entrapped in the sludge mass as it settles. This indicates that basin depth and detention are important in getting effluent SS down to low levels. Mixed liquor settling tests run at several treatment plants in Sweden (33) showed that residual turbidity above the sludge interface dropped significantly over the first hour.

7.6.2 Development of Loading Parameters from Mixed Liquor Settling Tests

7.6.2.1 Surface Hydraulic Loadings

Initial settling velocity (ISV) at actual mixed liquor concentration may be determined in a single test simply by plotting the height of the sludge-liquid interface vs. time and noting the slope of the straight line portion of the plot. The critical minimum ISV value for a particular system may be estimated from results of a number of individual tests. The designer should attempt to establish relations between ISV and biological process parameters such as mixed liquor concentration and organic loading. The selected ISV value should then reflect conditions most unfavorable to settling including correction for minimum expected temperature. Finally a capacity factor as discussed in Section 7.3.1 should be applied to convert the critical ISV to a hydraulic loading.

The resulting maximum surface hydraulic loading should not be exceeded by any sustained maximum flow (say 4-hr duration). Initial settling velocities for mixed liquor from air activated sludge systems have been reported to range from 3 ft/hr to over 20 ft/hr (4) (21) (34). For good settling (non-bulking) air activated sludges from municipal wastewaters the following design relation between the ISV and the mixed concentration has been suggested (34):

$$Vi = 22.5e^{-338Ci}$$

Where:

Vi = settling velocity in ft/hr
Ci = concentration in lb/lb

Bulking sludges will show ISV values well below those indicated by this line. Sludges with superior settling qualities may show considerably higher values.

7.6.2.2 Sludge Volume Index

The sludge volume index (SVI) widely used to guide operating control of the activated sludge process, provides an approximate indication of sludge compaction characteristics. The index is calculated by dividing the initial mixed liquor SS concentration (percent) into the settled volume (percent of initial volume) occupied by the solids after one half hour of settling.

The reciprocal of the SVI is often taken as an approximate indication of the maximum return sludge concentration which can be obtained with a given mixed liquor (100/SVI = percent solids). The index has been used as a guide to sizing return sludge pumping requirements to maintain different mixed liquor concentrations (2). Although the SVI does not give a direct indication of solids transport capacity, it has been suggested that for index values of less than 100, underflow concentrations below 1 percent and mixed liquor concentrations below 3000 mg/l, hydraulic rather than loadings will govern clarifier sizing (2).

7.6.2.3 Solids Loading

Based on the analysis discussed in Section 7.6.1, Dick (31) has proposed a method for determining limiting solids transport capacity as a function of underflow rate, given a curve or equation defining the relation of settling velocity to concentration. Dick and Young (35) have formulated the method into a series of equations, assuming that the settling velocity-concentration curve could be represented in the form:

$$V = a\,c^{-n}$$

where

V is settling velocity
c is concentration

and

a and n are appropriate constants for the units used.

The most serious problem in applying the method is determining the settling velocity-concentration relation. Dick suggests developing the relation from a series of ISV tests on the same mixed liquor at different initial concentrations (obtained by settling, decanting clear liquid and resuspending the solids). There is a serious question whether a curve developed from such tests really represents the behavior of the solids in the sludge blanket of a clarifier. Nevertheless this approach is the best presently available for estimating solids transport capacity fom settling tests. Others suggested (6) are open to even more serious objections.

In translating solids transport capacity to an allowable solids loading some safety factor may be needed to allow for possible critical conditions (temperature, poor thickening characteristics, etc.) not reflected in the test work.

In a design application trial solutions at different return sludge rates may be justified to determine the effect on tank sizing of the different solids loadings and capacities that result at the various underflow rates. Sizing should be based on peak solids loadings associated with sustained maximum flows unless specific testing has justified a reduction taking advantage of storage of peak solids by increases in sludge blanket height. Such storage should be avoided with nitrifying sludges (34).

7.6.2.4 Settling Test Procedures

Although it has been demonstrated (36) (37) that factors such as column diameter, sludge depth, dissolved oxygen and application of stirring can significantly affect the results of settling tests, standard values for such factors or standard allowances for their variations have not been adopted. Dick (36) has detailed test procedures and indicated (38) preference for use of sludge depths of 3 ft, column diameters of 3.5 in. or more and slow stirring at tip speeds of 10 in. min.

7.6.3 Flow Stabilization and Density Currents

Two approaches to preventing short circuiting from density currents, were described in Section 7.3.2: dynamic stabilization and density stabilization. An exhaustive study of shallow activated sludge settling tanks has been made in Sweden (33). Included were tracer studies on tanks under actual operating conditions and parallel quiescent and stirred settling column tests on the mixed liquor. These tanks, although designed for dynamic stabilization, showed serious short circuiting. Due to flocculation in the tanks, however, effluent quality was better than predicted from the quiescent settling results and actual detention times.

In the U.S. where final settling tanks commonly have design depths 10 ft or more, flow stabilization depends totally on density. Unfortunately studies of the type conducted in Sweden have apparently not been run on tanks designed for density stabilization. In the side-by-side performance tests (Section 7.4) comparing special inlet designs for circular tanks with conventional center feedwells, density stabilization could have been important, but no data were taken to show the degree of short circuiting. Neither were any parallel settling column tests run. Tracer studies of these special inlets generally have been run on clear water, so they fail to show any effects of density stabilization. Even without these effects, special inlets displayed less short circuiting (13) (14).

In an attempt to minimize undesirable density current effects, several designs have varied the placement of sludge drawoffs and effluent weirs in relation to the inlet. Sawyer (39) pointed out the "submerged waterfall effect" that occurs when the density current reaches the tank floor in its initial downward sweep. In conventional rectangular or circular tanks where the sludge drawoff is located below the inlet the impact of the "waterfall" can dilute the collected sludge and resuspend a portion of it. Peripheral-feed circular tanks avoid this problem as do those equipped with suction-type mechanisms which remove sludge from the entire tank floor. Even in tanks with centerfeed and center sludge drawoff, use of deep feedwells discharging at low velocities can minimize the problem (10). Rectangular tanks have been constructed with sludge drawoffs located away from the inlet. Excellent results have been obtained at New York City with sludge drawoffs at mid-length of the tanks (30). This arrangement uses the density current to speed sludge removal but prevents the density current from entering the outlet zone of the tank. Rectangular tanks may also be equipped with suction-type sludge removal on traveling-bridge mechanisms.

In one special rectangular tank arrangement, effluent weirs are distributed over the length of the tank, with baffles in the upper part of the tank to impede counter currents induced by density current below and force vertical flow to the weirs. In peripheral feed circular tanks, tests have shown that units with peripheral drawoffs located just inside the inlet channel produce better effluent than units with weirs located more toward the tank center (40).

Weir hydraulic loadings of 15,000 gpd/ft at average design flows are suggested in the Ten State Standards (29), with allowances of up to 20,000 gpd/ft where weirs are located so that density currents do not upturn below them. Loadings of up to 100,000 gpd/ft have been used without apparent problems in designs such as those of New York City where weirs are well separated from density current effects (30).

7.6.4 Sludge and Skimmings Removal

Suction-type sludge removal should be considered wherever sludge detention in the tanks is critical and doubt exists about conveyance time for other mechanisms. Desirable features in suction-type mechanisms include independent flow controls for each suction drawoff and visible gravity sludge discharges.

Federal guidelines (41) require skimming equipment on secondary settling tanks to remove floating sludge and any oily materials not separated in previous treatment. Scum quantities generally are small in relation to those from primary tanks (0.1 cu ft mg). Where no primary tanks are included in plant process, scum quantities from secondary tanks could be conservatively estimated on the same basis as for primary tanks. Effluent weirs should be laid out to permit skimming the maximum possible portion of the tank surface.

Maximum practical concentrations of underflow from secondary clairifiers in activated sludge systems range from 0.5 to 2.0 percent solids, depending on settling and compaction characteristics of the sludge. Actual concentrations depend on the return sludge pumping rate. Sludge concentrations of 3 to 7 percent solids may be obtained from secondary clarifiers in trickling filter systems.

7.7 Chemical Sedimentation

Sedimentation of chemically coagulated or precipitated wastewaters is similar to sedimentation of wastewaters without chemicals. The design of tanks can proceed on essentially the same basis, provided special consideration is given to the effects of chemical treatment on settling characteristics, sludge quantities, resistance of the sludge to movement by collecting and pumping equipment, and the special maintenance problems encountered with lime coagulation. Few data have been reported concerning settling characteristics of chemically precipitated floc in wastewater treatment. Some data are available on chemical precipitation in water treatment using similar chemicals (42).

From the literature it is apparent that actual surface loadings vary considerably from one application to another (2) (43) (44) (45) (46) (47) (48) (49). This wide variation emphasizes the importance of testing and pilot work in designing chemical precipitation facilities. In the absence of testing indicating higher figures to be satisfactory, the following typical surface loading rates may be used for sizing tanks (47) (50):

Chemical	Peak Surface Loading gpd sq ft
Alum	500-600
Iron	700-800
Lime	1400-1600

In general, these design rates may be used for primary, secondary or tertiary applications. It should be noted, however, that they are based on limited data and may be revised when more experience is available.

Sludge quantities from chemical precipitation can be estimated from the SS removal and the stoichiometry of chemical reactions involved. Volumes depend on sludge concentrations which are highly variable (1 to 15 percent) and are best determined by actual test. Equipment suppliers should be consulted about strength and power of collector equipment to handle the dense sludges expected from lime precipitation. Extra smooth piping glass-lined or PVC, should be used for lime sludges. Average sludge productions determined from raw wastewater coagulation by lime, iron and alum are 6,500, 1740 and 1120 lb mg respectively (47). Average sludge volumes for the same locations for lime and iron are 10,000 and 13,000 gal mg, respectively (47). Brown (51) observed a sludge production of 1894 lb solids mg (6275 gal mg) using alum for precipitating trickling filter effluent.

7.8 Flotation

7.8.1 Applications

This section deals with flotation induced by introduction of fine gas bubbles into wastewater. Since most SS in municipal wastewater have specific gravity values only slightly above 1, adhesion of the gas bubbles to the solids particles readily makes them buoyant.

For flotation of solids in municipal wastewater, gas bubbles must be quite fine (.01 to 0.1 mm); otherwise, their own rise rate prevents significant adhesion to the solid particles. Three methods of introducing gas bubbles have been shown to create bubbles sufficiently fine for flotation of municipal wastewater SS. Vacuum flotation and dissolved-air flotation (DAF) both create conditions in which the wastewater is supersaturated with air at some pressure. Upon reduction of that pressure, air comes out of solution as finely-divided bubbles. Auto-flotation can occur in algae suspensions if they become sufficiently supersaturated with dissolved oxygen from photosynthesis. Vacuum flotation and autoflotation are not often used because the former is expensive and the latter can only operate under limited conditions of warm weather and bright sunshine (52). Diffused or submerged turbine aerators create bubbles much too coarse for flotation of municipal wastewater solids.

Pressure and vacuum flotation units have found only limited application in treatment of municipal wastewater. It has been difficult to justify using these units in conventional applications such as primary SS removal or mixed liquor clarification because sedimentation is ordinarily cheaper, simpler and often provides better results.

Advantages which might favor use of flotation in special applications include: 1) higher surface loadings, hence smaller tanks sizes (important where space is critical); 2) ability to handle peak seasonal loads or storm flows (in some designs flotation may be used intermittently to increase capacity of settling tanks); 3) effectiveness in removing solids which are difficult to settle.

Dissolved-air flotation has been suggested for separation of grit and scum in a single treatment unit (2) (53) (54). Performance data for such applications are lacking, however. Because it can produce a float of much thicker consistency than settled activated sludge, dissolved-air flotation has been tried for mixed liquor solids separation. Full scale studies at

Manassas, Va. (55) indicated that this application was not economically competitive with conventional systems.

7.8.2 Dissolved-Air Flotation

As shown in Figure 7-7, dissolved air flotation (DAF) units commonly employ rectangular tanks with separate chain-and-flight scum and sludge collectors. Circular units are also commercially available. The widest application for these units has been as thickners for waste activated sludge. Units used for SS separation are similar, but design parameter values vary according to the application. To avoid fouling of pressurizing and pressure-regulating equipment and excessive shearing of influent solids, a stream of recycled effluent is usually pressurized. Upon pressure release, this stream is blended with the inflow to be treated. Other methods include pressurizing all or part of the influent stream.

Design of DAF units involves selection of values for a number of parameters including percent recycle flow, operating pressure, pressurization retention time, air flow, and surface hydraulic loading, solids loading (area basis) and float detention period. Variables reflecting influent characteristics include flow, solids loading, liquid temperature and type and quality of influent solids. Investigators have attempted to relate flotation performance to the air to solids ratio and a number of other variables with a limited amount of success.

Mulbarger and Huffman (56) noted that float concentrations depend more on float detention time than solids loading. They related capture in flotation to a parameter equal to the air to solids ratio divided by the product of surface hydraulic loading and dynamic viscosity.

Values of specific parameters used in actual applications vary widely. Typical ranges cited are as follows (2) (50) (56) (57) (58) (59) (60) (61):

Parameter:	Range:
Pressure, psig	25 to 70
Air to Solids Ratio, lb/lb	0.01 to 0.1
Float Detention, min	20 to 60
Max. 24-hr	
Surface Hydraulic Loading, gpd/sq ft	500 to 4000
Recycle, percent	5 to 120

Available data from specific applications are summarized in Table 7-4.

In flotation equipment special attention must be given to the inlet, outlet and collector mechanism configurations. The flotation tank must permit aggregate rise with a minimum of interference in the form of turbulence or obstructions and provide for removal of floated froth, settled sludge and treated effluent. Effluent ports must be sufficiently submerged to prevent interference with the froth on the surface. The inlet conditions of the flotation tank are critical to proper performance. Baffles, walls, and other obstructive energy-dissipating devices tend to destroy aggregate bonding with resulting loss in flotation efficiency. Also,

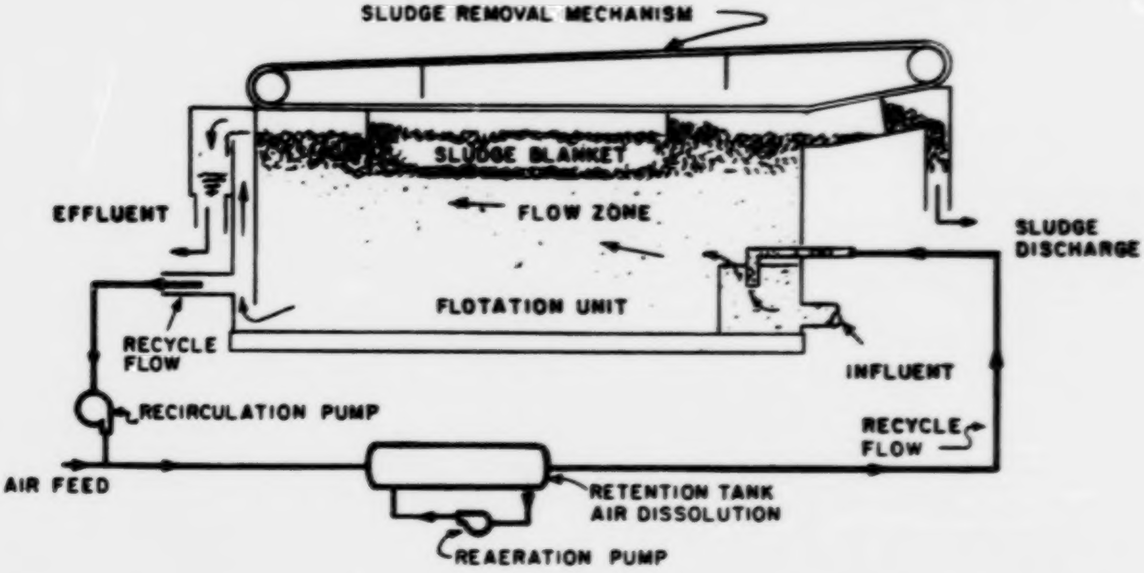

SLUDGE REMOVAL MECHANISM
SLUDGE BLANKET
EFFLUENT
FLOW ZONE
SLUDGE DISCHARGE
FLOTATION UNIT
RECYCLE FLOW
INFLUENT
RECIRCULATION PUMP
RECYCLE FLOW
AIR FEED
RETENTION TANK AIR DISSOLUTION
REAERATION PUMP

FIGURE 7-7
SCHEMATIC OF A DISSOLVED-AIR FLOTATION UNIT
(Courtesy of Komline-Sanderson)

TABLE 7-4

DISSOLVED-AIR FLOTATION APPLICATIONS

			Chemical Treatment		Flotation		SS Performance Data				
Plant	Design Flow	Type of Municipal Wastewater	Coagulant	Dose	Surface Hydraulic Loading	Detention Time	Inf.	Eff.	Removal	Ref.	Remarks
	mgd			mg/1	gpd/sq ft	hrs.	mg/1	mg/1	percent		
Aker, Sweden	0.4	Primary Effluent	Alum(a)	100	2360	0.31	–	–	71(c)	(62)	
Klagerup, Sweden	0.12	Primary Effluent	Alum(a)	159	1180	0.37	–	–	77(c)	(62)	
Salemstaden, Sweden	2.16	Primary Effluent	Alum(a)	175	2540	0.26	–	–	60(c)	(62)	
Bara, Sweden	0.08	Aerator Mixed Liquor	Alum	–	2480	0.24	–	–	–	(62)	
Kungsor, Sweden	1.93	Unsettled Trickling Filter Effluent	Alum(a)	145	4480	0.20	–	–	97(d)	(62)	
Flen, Sweden	2.54	Unsettled Trickling Filter Effluent	Alum(a)	–	3360	0.35	–	–	–	(62)	
Prince William County, Va.	1.0	Aerator Mixed Liquor	None	–	360	3.4	–	30 to 100	–	(56)	Limiting solids loading 15 lb/lb SF
Bellair, Texas	(b)	Aerator Mixed Liquor	Cationic Polymer	8 30		–	2000	70 17		(63)	
Stockton, Calif.	(b)	Lagoon Effluent	Alum	75-225		0.32	94 to 152	12 to 20	87	(52)	Includes filtration

(a) 30-60 min. of flocculation provided before flotation. (b) Pilot Plants.

(c) BOD removal; no SS data given. (d) P removal; no SS data given.

turbulence in the region of the froth will result in losses of floated solids. Ettelt (58) reported several different designs of inlet structures in his prototype units. His tangential flow inlet appears to offer considerable promise where such designs are compatible with the entire structure.

Further discussion of design features of dissolved-air flotation units may be found in the literature (56) (58) (59) (60).

7.9 Shallow Settling Devices

The potential advantages of multiple tray shallow settling devices have long been recognized (3) (4), but early prototypes of such equipment were unsuccessful due to practical problems of flow distribution and sludge removal. In recent years, shallow settling devices of improved design, such as tube settlers, have been applied to water and wastewater treatment. Tube settlers consist of bundles of small plastic tubes with hydraulic radii ranging from one inch upward and lengths of 2 ft or more, depending upon the particular application. Square tube sections are most common but hexagonal and other shapes have been used by various manufacturers.

Tubes are commonly inclined steeply (60°) to horizontal and fabricated in modules, as shown in Figure 7-8. These modules have beam strength which permits their installation in settling tanks, as shown in Figures 7-9 and 7-10. Clarifier influent is introduced beneath the tube modules. The flow passes upward through the modules with the solids moving countercurrently by gravity (Figure 7-11) and falling from the tube bottoms into the sludge collection zone beneath. The clarified effluent is collected above the tube modules.

Free standing package units with tubes only slightly inclined (5°) have found some application in small chemical clarification filtration systems for tertiary wastewater treatment.

Tube settlers promote sedimentation in three ways: 1) the multiple tubes stacked one above another provide an effective settling area several times that of the projection in plan of the modules; 2) the small hydraulic radius of the tubes maintains laminar flow and promotes uniform flow distribution; 3) in steeply inclined tubes, the movement of sludge against the direction of flow favors particle contact and agglomeration. This additional flocculation offsets the reduction in their horizontal projected area caused by inclining the tubes. For alum floc suspensions a given length tube has been shown to provide most effective removal at an inclination of about 45 degrees, and performance even at 60 degrees was comparable to that when horizontal (64).

Tube settlers have been promoted both for reducing required size of settling tanks and for improving their performance, but manufacturers presently tend to emphasize improved performance and recommend the same surface hydraulic loadings for tanks equipped with tube settlers as for conventional tanks.

Comparative data on performance of tanks with and without tube settlers (either side-by-side or before-and-after) are shown in Table 7-5. The data are quite limited and

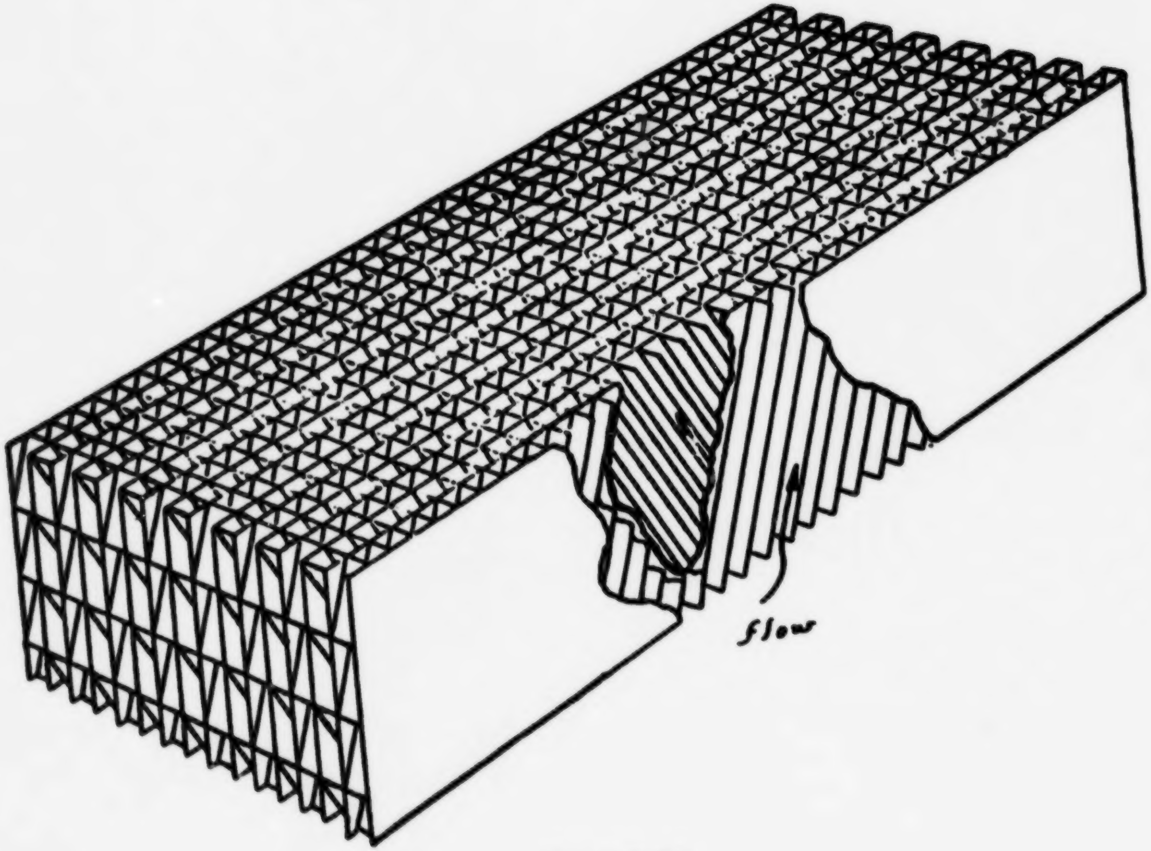


FIGURE 7-8

MODULE OF STEEPLY INCLINED TUBES

(Courtesy Neptune Microfloc, Inc.)

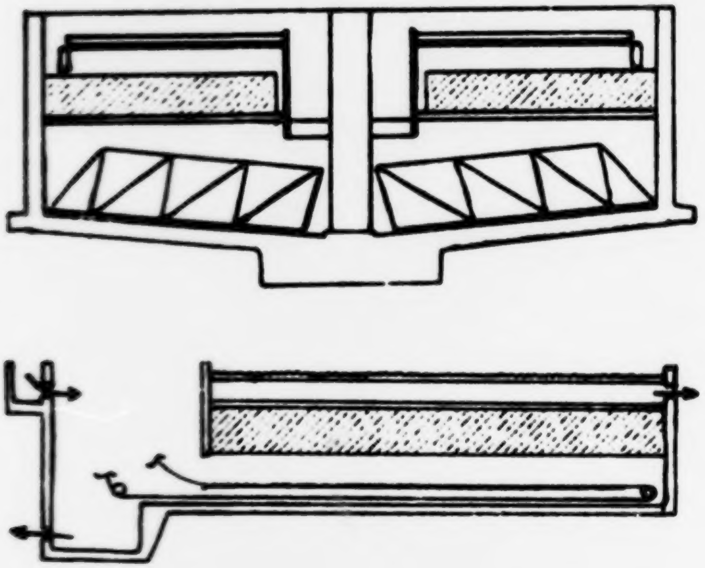


FIGURE 7-9
TUBE SETTLERS IN EXISTING CLARIFIER

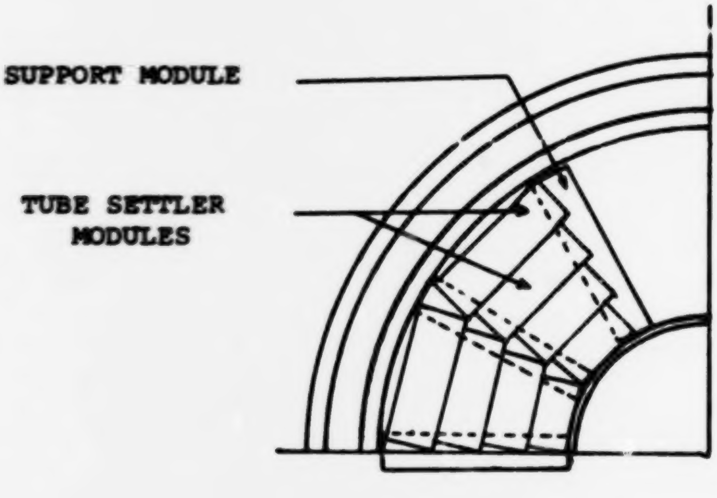

SUPPORT MODULE
TUBE SETTLER MODULES

FIGURE 7-10
PLAN VIEW OF MODIFIED CLARIFIER

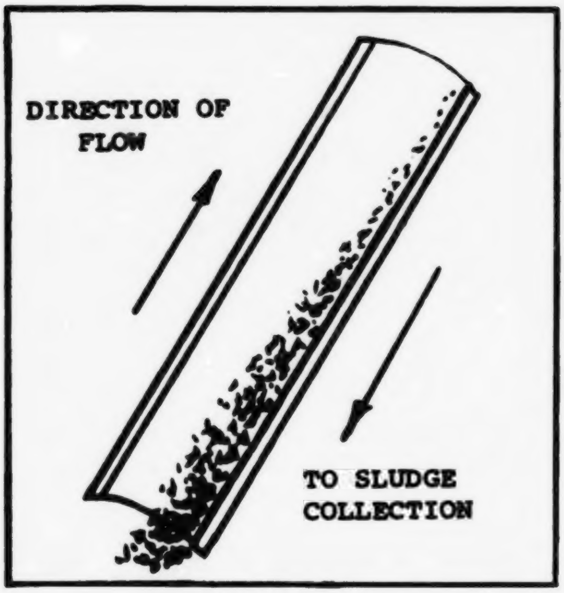

DIRECTION OF FLOW
TO SLUDGE COLLECTION

FIGURE 7-11

TUBE SETTLERS – FLOW PATTERN

TABLE 7-5

TUBE SETTLER INSTALLATIONS

		Plant Flow			Existing Facility		Operational Data Using Tube Settlers			
Plant Location	Type	Design mgd	Actual mgd	Location	Overflow Rate gpm/sq ft	Effluent SS mg/l	Tube Overflow Rate gpm/sq ft	Tank Overflow Rate gpm/sq ft	Effluent SS mg/l	Reference
Hopewell Township,	Activated	0.13	0.13	Secondary	0.34	60-70	–	–	–	65
Pennsylvania	Sludge			Clarifier	–	–	2	0.68	27	
Trenton,	Activated	6.5	5.6	Secondary	–	–	0.56	0.29	8	66
Michigan	Sludge			Clarifier						
Lebanon,	Activated	0.75	1.25	Secondary	0.61	61	0.85	0.61	30	67
Ohio	Sludge			Clarifier						

rather inconclusive as to the benefits obtained from the tube settlers.

Tube settlers have found wider application in water treatment than in wastewater. For wastewater, tube settlers may find their best applications in tertiary coagulation and settling. They also may be of help in upgrading performance of units with serious short circuiting problems.

When installed, settling tubes usually cover one half to two thirds of the basin area. To prevent fouling of tubes, the remaining area between the inlet and tube area is arranged to provide scum removal. The portion of the basin equipped with settlers should have collecting weirs at 15 ft or closer spacing to induce an even vertical flow distribution and reduce short circuiting.

Tube settler installations require a support grid (usually designed to support one man) and surface baffles to separate the tube settler and scum collection area. Minimum basin depths should be 10 to 12 ft.

Long term studies have revealed that in wastewater treatment the upper surface of the settling tubes becomes coated with sludge (68). Long term fouling of the tubes with grease or rags has not been a problem, but in order to maintain a high degree of solids removal, it is generally necessary to install an air grid and periodically interrupt flow to introduce air to remove the sludge build-up on the tubes.

A typical air wash cycle consists of draining the tank to the level of the tubes and then allowing air to rise up through the tubes. The air is supplied either from a fixed grid or a scour system attached to the rake arm. After the air wash, approximately 15 to 25 min is required for the effluent SS to return to normal, e.g., from 60 mg l to 10 mg l before the unit can be returned to service (68). A short quiescent period of no flow may also be needed between the drain down and the air wash (69). Generally, required cleaning frequency varies from one week to several months (70). Where serious sludge carryover conditions are experienced, however, it has proven difficult to prevent fouling with even the highest cleaning frequency (71).

7.10 Wedge-Wire Settler

Wedge-wire settlers are wire matrices installed in clarification basins similarly to tube settlers. They are designed to improve the quality of settled effluent from activated sludge or trickling filter treatment.

The equipment can be installed in any conventional clarifier configuration. The settling device consists of a matrix constructed from parallel wires (See Figure 7-12) suspended beneath and parallel with the surface of the water so that the wastewater must pass upward through the mesh before reaching the effluent weir. Over 200 secondary clarifier installations in England are equipped with wedge-wire settlers (72).

Typically, the wire in the matrix is triangular in cross-section and arranged with apex point-

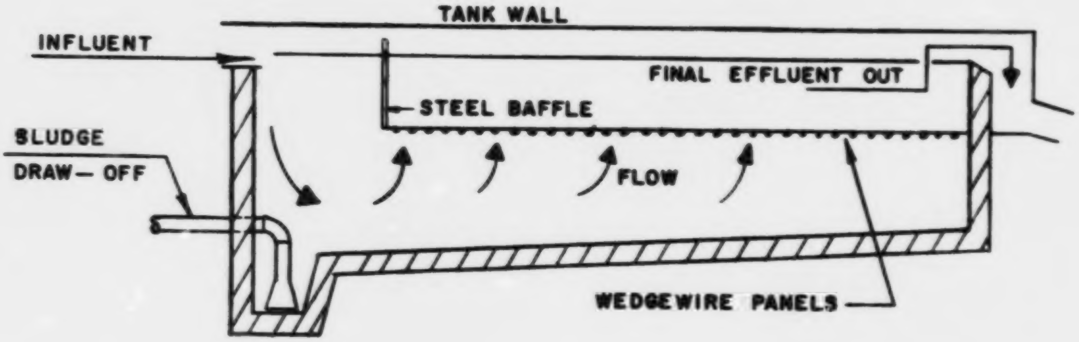

TANK WALL
INFLUENT
FINAL EFFLUENT OUT
STEEL BAFFLE
SLUDGE DRAW— OFF
FLOW
WEDGEWIRE PANELS

FIGURE 7-12

SIMPLE WEDGE WIRE CLARIFIER

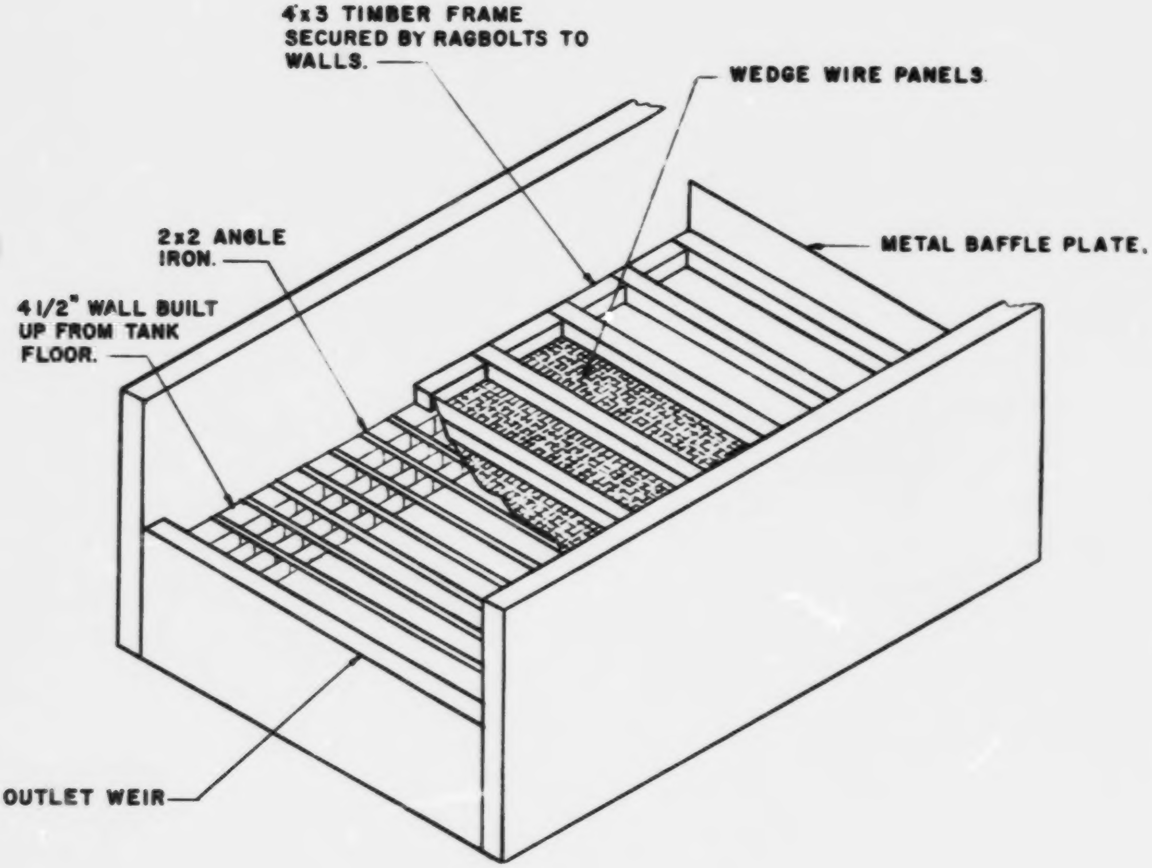

4'x3 TIMBER FRAME SECURED BY RAGBOLTS TO WALLS.
WEDGE WIRE PANELS.
METAL BAFFLE PLATE.
2x2 ANGLE IRON.
4 1/2" WALL BUILT UP FROM TANK FLOOR.
OUTLET WEIR

FIGURE 7-13

INSTALLATION OF WEDGE WIRE PANELS

ed downward, to provide 0.125 to 0.250 mm openings at the top surface. The openings comprise about 15 percent of the total area. Construction is of either stainless steel or aluminum and the wires must be rigidly fixed. The wire "rack" is supported within the tank about 6-inches below the water surface on steel angles or other similar structural grid-work (See Figure 7-13).

Used in conjunction with effluent launders at spacings of 15 ft or less, wedge-wire settlers distribute flow quite uniformly over the entire area of the basin, producing a nearly ideal upflow clarification zone above the wire (65). Finer particles which settle in this zone eventually coalesce into a sludge blanket which aids in removal of near-colloidal particles. When the blanket builds in thickness and nears the water surface (2 to 4 days) cleaning is necessary. The basin level is lowered below the wire level and the wire is hosed down to clean off the accumulated solids. Total clean up time is about one hour (72).

Wedge-wire settler applications have been limited to relatively low surface hydraulic loadings, i.e., 600 gpd/sq ft with peaks to 800 gpd/sq ft. Effluent quality has been roughly related to flow rate (73). Under stable conditions an effluent SS of 15 to 20 mg/l could be expected at 600 gpd/sq ft and 10 mg/l at 300 gpd/sq ft. Results of side by side tests of secondary clarifiers treating trickling filter effluent (150 mg/l SS) indicated that standard clarifiers produced effluents from 7 to 77 mg/l with an average of 41 mg/l while identical wedge-wire settlers produced effluents of 1.6 to 18 mg/l with an average of 8 mg/l (74). All units operated at a relatively low rate of 300 gpd/sq ft. This low hydraulic loading appeared significant to the wedge-wire settler performance, but it had little effect on the effluent quality of the conventional settling unit.

In activated sludge clarification use of the wedge-wire settlers reduced effluent SS to 8 to 16 mg/l compared with 30 to 40 mg/l from the same conventional clarifier at a rate of 600 gpd/sq ft (74).

Pullen (75) cited 5 clarifiers of small size (18,000 to 170,000 gpd), which were equipped with wedge-wire screens, experiencing a 50 percent improvement in effluent SS quality.

Wedge-wire settlers are limited to multiple tank installations so that during shut down for washing flow can be diverted to other clarifiers. Drain and wash down flow can be recycled to pretreatment units or to sludge handling systems.

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CHAPTER 8

PHYSICAL STRAINING PROCESSES

8.1 General

Physical straining processes are defined for the purpose of this manual as those processes which remove solids by virtue of physical restrictions on a media which has no appreciable thickness in the direction of liquid flow.

Physical straining devices may be grouped according to the nature of their straining action. (See Table 8-1).

8.2 Wedge Wire Screens

8.2.1 Inclined Screens

Inclined screens, are typified by the Hydrasieve, (Figure 8-1) made by C-E Bauer, Division of Combustion Engineering Inc., or the Hydroscreen made by Hydrocyclonics Corporation. These devices were originally developed in 1965 for the pulp and paper industry to dewater and classify pulp slurries having solids contents of 6 percent or less (1). The units operate by gravity and function as an inclined drainage board with a screen of wedge wire construction having openings running transverse to the flow.

The first full scale municipal application of Hydrasieves was at the Ohio Suburban Wastewater Treatment Plant at Huber Heights in 1967 treating raw wastewater (1).

8.2.1.1 Equipment Details

The screen consists of three sections with successively flatter slopes on the lower sections (Figure 8-2). The screen wires are triangular in cross section as shown in Figure 8-3, and usually spaced 0.06 in. apart for raw wastewater screening applications. In the Bauer unit, these wires bend in the plane of the screen, as illustrated in Figure 8-4. They are straight and transverse to the flow in the Hydrocyclonics unit.

Above the screen and running across its width is a headbox; Figure 8-2 shows two possible inlet designs. A light-weight hinged baffle at the top portion of the screen reduces flow turbulence in the Bauer unit. To collect the solids coming off the end of the screen several arrangements can be used, including a trough with a screw conveyor.

Inclined screening units are generally constructed entirely of stainless steel. Lighter units with a fiber glass housing and frame costing about 25 percent less (1) may also be obtained. Dimensions and capacities for hydrasieve units are given in Table 8-2.

TABLE 8-1

PHYSICAL STRAINING PROCESSES

Principal Applications	Device	Hydraulic Capacity	Straining Surface	Waste Solids Composition	Percent SS Removals
Pretreatment & Primary Treatment	Inclined wedge-wire stainless steel screens	High flow rates 4-16 gpm/in of screen width	Coarse .01 to .06 in (250-1500 microns)	10 to 15% solids by weight	5-25
" "	Rotary stainless steel wedge wire screens	16-112 gpm/sq ft	Coarse .01 to .06 in (250-1500 microns)	16 to 25% solids by weight	5-25
" "	Centrifugal screens	40-100 gpm/sq ft	Medium 105	0.05-0.1%	60-70
Secondary and Tertiary SS Removal	Micro-Screens	Medium flow rates 3 to 10 gpm/sq ft	Medium[b] 15-60 microns	[illegible] mg/l (app. 0.05%)	50 to 80
" "	Diatomite filters	Medium flow rates 0.5-1.0 gpm/sq ft	N/A[b]		up to 90
" "	Ultra-Filters	Low flow rates 5 to 50 gpd/sq ft	Fine[a] 10^{-3}-15 microns	1500 mg/l (1.5%)	> 99

(a) These values typify the range of solids filtered by the media. Removals are a function of media thickness and not media opening sizes.

(b) Straining occurs through particulate mat of solids on screening surface.

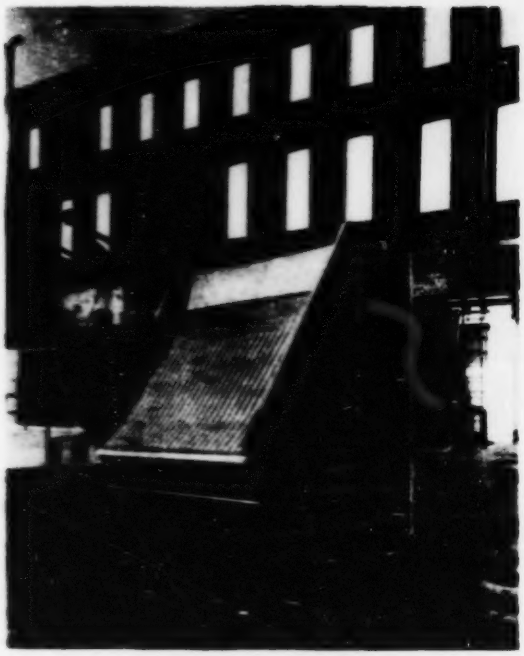

FIGURE 8-1

HYDRASIEVE SCREENING UNIT

Courtesy C-E Bauer

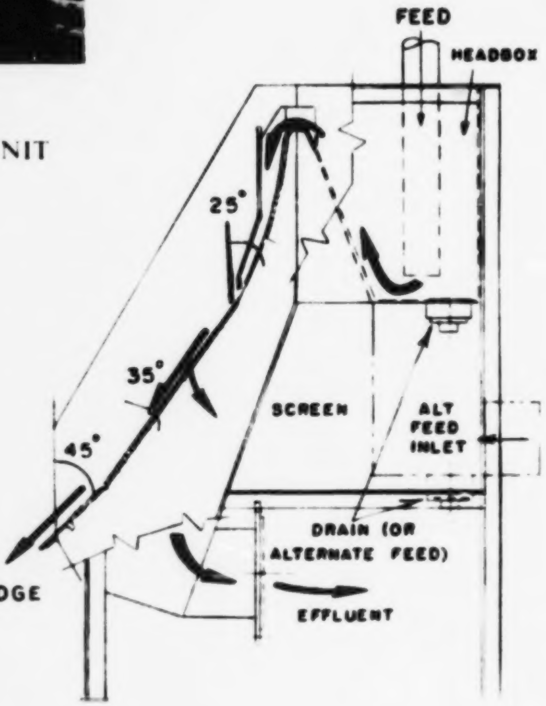
FEED
HEADBOX
25°
35°
45°
SCREEN
ALT FEED INLET
DRAIN (OR ALTERNATE FEED)
SLUDGE
EFFLUENT

FIGURE 8-2

HYDRASIEVE SCHEMATIC

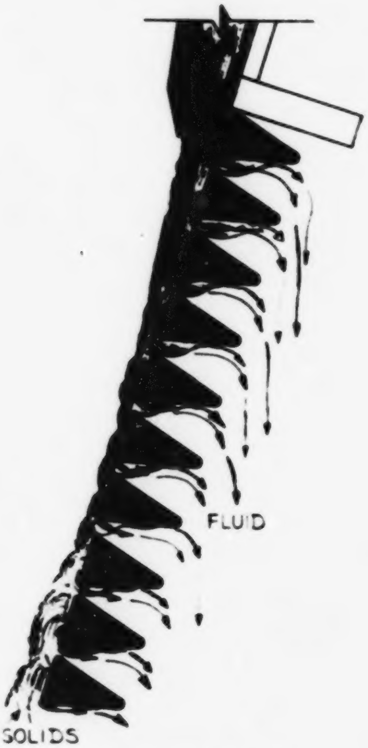
FLUID
SOLIDS

FIGURE 8-3

SCREEN DETAIL

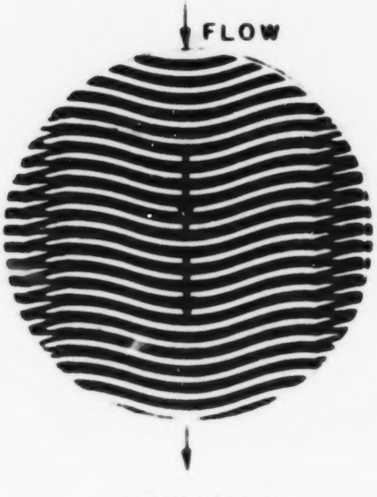

FIGURE 8-4

CURVED SCREEN BARS

Courtesy C-E Bauer

TABLE 8-2

SPECIFICATIONS OF HYDRASIEVES

OVERALL DIMENSIONS

Width ft	Depth ft	Height ft	Weight lb	Capacity mgd
2	3.5	5	350	0.2
3.5	4	5	550	0.4
4.5	5	7	650	0.9
5.5	5	7	800	1.2
6.5	5	7	1000	1.5
7	9.5	7.3	1800	2.9
14	9.5	7.3	3600	5.8
21	9.5	7.3	5400	8.7
28	9.5	7.3	7200	11.6
35	9.5	7.3	9000	14.5

8.2.1.2 Process Description and Design

Influent wastewater enters and overflows the headbox, on to the upper portion of the screen. On the screen's upper slope most of the fluid is removed from the influent. The solids mass on the following slope, because it is flatter, and additional drainage occurs. On the screen's final slope the solids stop momentarily, simple drainage occurs, and the solids are displaced from the screen by oncoming solids (2).

In test studies and actual installations hydrasieves have been operated satisfactorily at loading capacities of 4 to 16 gpm per in. of screen width (1). This hydraulic capacity is a function of the viscosity (which is a function of the temperature of the fluid), the solids loading, and the spacing of the individual slots. Slot width is selected by actual tests using sample screens. Once the slot opening has been chosen the screen's capacity per foot of width can be determined from empirical relationships. Since work to date has not been sufficiently extended to actual municipal wastewater conditions, pilot studies should be the prime basis for design.

Little quantitative work has been done on the solids loading capacity of a hydrasieve but generally speaking, for good performance, the influent should be dilute enough for smooth flow over the weir. Unit sizes designed to accommodate more than 1 mgd are available; however, for pilot studies a 6-in. wide by 22 in. long screen can be used provided flow rates are limited to 5 to 10 gpm (1).

8.2.1.3 Operating Experience

At the 3 mgd Huber Heights plant in Ohio hydrasieves ahead of trickling filters have effectively replaced primary clarifiers. Using 3 hydrasieves 72 in. wide and 54 in. long with a slot opening of 0.06 in., an average suspended solids removal of 25 percent was obtained while the units operated over a flow range of 1.5 to 4.5 mgd. Roughly 1 cubic yard of solids was removed per million gallons of wastewater with an average solids content of 12 to 15 percent (3)(4).

Although inclined screens cannot remove SS to the same extent as a sedimentation tank, they have been favorably received by operators because they do an excellent job of removing trashy materials which may foul subsequent treatment of sludge handling units. Their ability to remove fine grit is limited by size openings. Separate grit removal equipment, if needed, should be installed after the inclined screens. In a pilot study at South Buffalo Creek Sewage Treatment Plant at Greensboro, North Carolina, hydrasieve suspended solids removals ranged from 10 to 30 percent with an average removal of 20 percent (5).

At the U.S. EPA Blue Plains pilot study in Washington, D.C. (6) hydrasieves were installed in an effort to eliminate operational problems of debris collection on the mixers and plugging of recycle and waste discharge lines. Although the screens eliminated these problems, suspended solids removals varied from only 7 to 11 percent. The low removals were attributed to the wastewater's age (24 to 48 hours) (7).

An installation list is included (Table 8-3). Some of these installations are temporary; hydrasieves are being used for short term alleviation of excess solids and flows coming into plants which are to be abandoned when new facilities are built.

Operating experience in these installations varies as to cleaning and maintenance requirements. Generally, a daily washing of the screen surfaces, which takes about 5 minutes, is sufficient for good screen performance. Washing is normally done with steam or hot water to remove grease which accumulates and blinds the screen preventing passage of wastewater through the screen, and resulting in poor separation (5).

Daily steam cleaning proved necessary at Freehold, N.J. (8) but other installations such as Huber Heights required only monthly steam cleaning (3). Grease build-up requiring steam cleaning appears to be related to low air and wastewater temperatures, exposure of units and high grease content in wastewater.

Incidental to the removal of suspended solids in this process is the aeration of the separated water. At the Huber Heights plant raw wastewater impinging on a Bauer screen has been found to be aerated up to a level of 2 or more mg/l of dissolved oxygen (3). A noticeable reduction in odors from the grit removed in the subsequent chamber has also been claimed along with the elimination of scum in the digester (3).

8.2.2 Rotating Wedge Wire Screens

Hydrocyclonics, to overcome grease blinding problems of its own wedge wire screen, developed a rotating wedge wire screen which backwashes itself (Figure 8-5). Wastewater passes vertically downward from the outside to the inside of the drum by gravity. The screened wastewater then passes out through the lower half of the drum to a collection trough.

TABLE 8-3

WEDGE WIRE SCREENS MUNICIPAL TREATMENT INSTALLATIONS

All units below are Bauer units unless otherwise indicated:

PLANT & LOCATION	REMARKS
Ohio Suburban Water Co. Huber Heights, Ohio	3-4 mgd
Rochelle Treatment Plant Rochelle, Illinois	
Prophetstown Treatment Plant Prophetstown, Illinois	
Corinna Treatment Plant Corinna, Maine	
Bucks County Sewage Authority Mr. MacNamara Executive Director (Hydrocyclonics Units)	
Upper Gwynedd Towamencin Municipal Authority Lansdale, Pa.	
Irvine Ranch Water District Irvine, California	
Hercules—AWT Div. Freehold Township, N.J. Mr. Ron Lee, Engineer	
Montgomery County Commissioners Moraine, Ohio	10 mgd
STP Rogersville, Tennessee (Hydrocyclonics Units)	
Blue Plains Pilot Study Washington, D.C.	
S. Buffalo Creek STP Greensboro, N. Carolina	0.03 mgd

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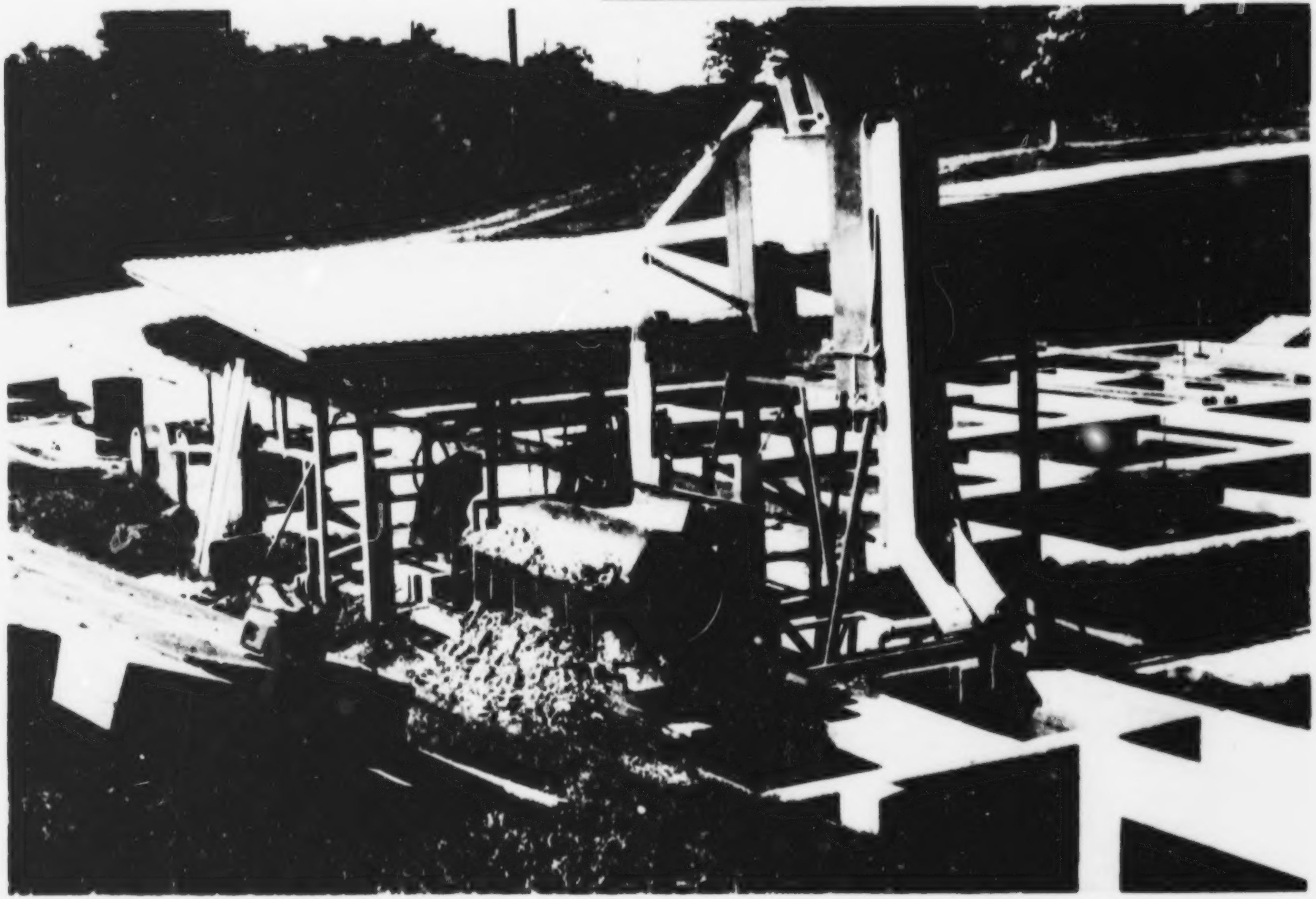


FIGURE 8-5

ROTATING WEDGE WIRE SCREEN AT NORTH CHICAGO S.T.P.
(Courtesy of Hydrocyclonics Corp.)

Solids are retained on the outside of the drum and are removed by a fixed scraper blade. A screen spacing of 0.06 inches is recommended for service on raw wastewater. In comparison with static screens, the manufacturer claims the rotating units require less maintenance, lower operating head and smaller space and produce dryer solids (9). Table 8-4 shows comparative design data for rotating and stationary wedge wire screens. (10).

TABLE 8-4

DATA SHEET-WEDGE WIRE SCREENS

Parameter	Inclined	Rotary
S.S. Removal, percent	5 to 25	5 to 25
Flow rate	4 to 16 gpm per inch of screen width	15 to 112 gpm sq ft
Wire Spacing	.01 to .06 in.	.01 to .06 in.
Percent solids by wt.	12 to 15	16 to 25
Volume of solids produced	1 to 2 cu yd of solids per million gallons of wastewater	

8.3 Microscreening

8.3.1 General Description

As shown in Figure 8-6 in its usual configuration a microscreen unit consists of a motor driven rotating drum mounted horizontally in a rectangular chamber. A fine screening media covers the periphery of the drum. Feedwater enters the drum interior through the open end and passes radially through the screen with accompanying deposition of solids on the inner surface of the screen. At the top of the drum pressure jets of effluent water are directed onto the screen to remove the mat of deposited solids. The dislodged solids together with that portion of the backwash stream which penetrates the screen are captured in a waste hopper as shown in Figure 8-7. Solids flushed from the unit are sent to sludge handling systems or recycled to the head of the plant. Units may be equipped with ultraviolet lights to control biological growth on the screen media. Effluent passes from the chamber over control weirs oriented perpendicular to the drum axis.

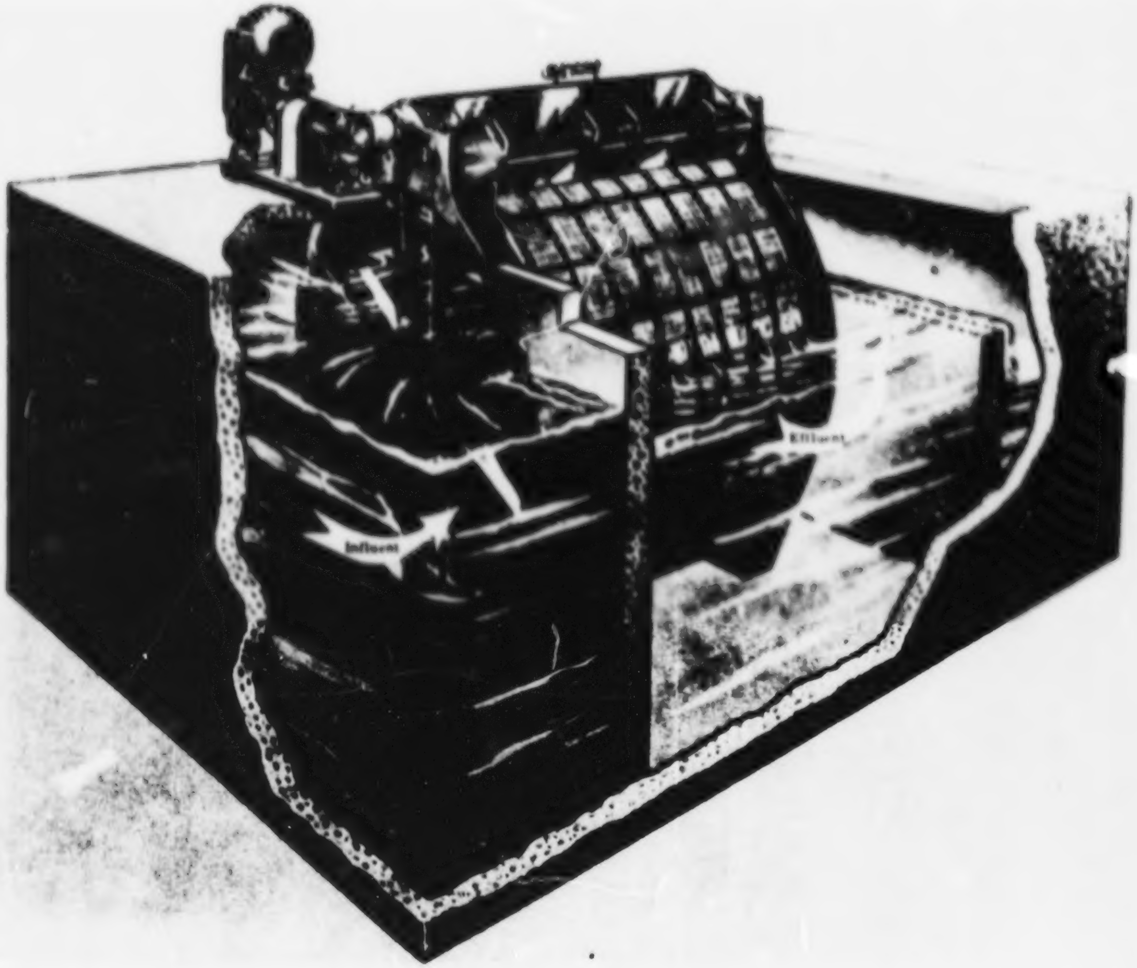

Influent
Effluent

FIGURE 8-6
TYPICAL MICROSCREEN UNIT
(Courtesy of Cochrane Division, Crane Co.)

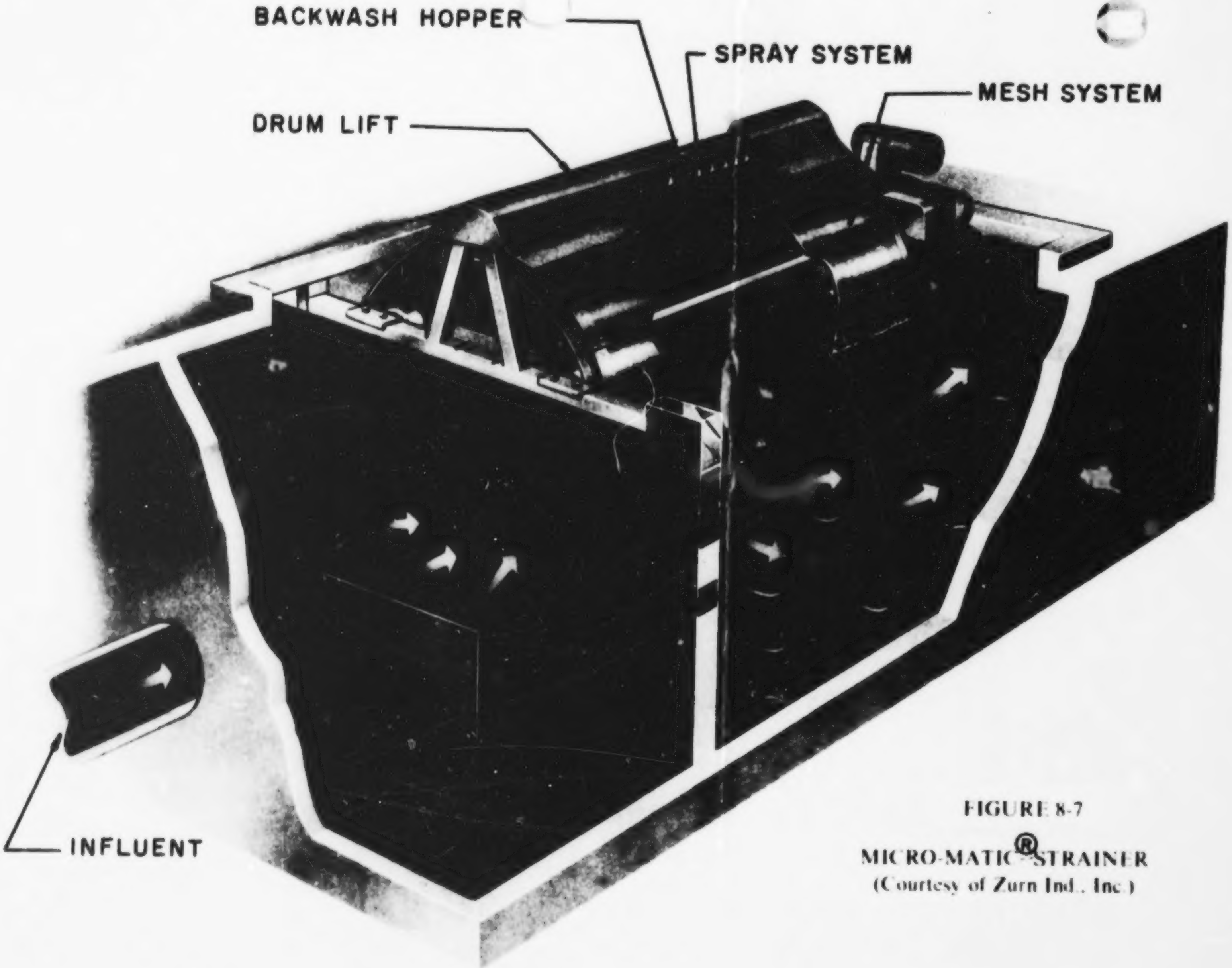

BACKWASH HOPPER
SPRAY SYSTEM
MESH SYSTEM
DRUM LIFT
INFLUENT

FIGURE 8-7

MICRO-MATIC® STRAINER
(Courtesy of Zurn Ind., Inc.)

8.3.2 Functional Design

The functional design of a microscreen unit involves:

1. Characterization of suspended solids in feed as to concentration and degree of flocculation, as these factors have been shown to affect microscreen capacity, performance and backwashing requirements (9)(11)
2. Selection of unit design parameters which will assure sufficient capacity to meet maximum hydraulic loadings with critical solids characteristics, and provide the required performance over the expected range of hydraulic loadings and solids characteristics.
3. Provision of backwash and supplimental cleaning facilities to maintain the design capacity.

Table 8-5 shows typical values for microscreen and backwash design parameters for solids removal from secondary effluents. Similar values would apply to direct microscreening of good quality effluent from fixed film biological reactors such as trickling filters or rotating biological contactors, where the microscreens replace secondary settling tanks (9). This application is not widely practiced, however.

Microscreening has been used for the removal of algae from uncoagulated lagoon effluents. At Bristol, England, algae reductions of 1565 to 450 algae per ml and 989 to 168 algae per ml were achieved on astrerionella, cyclolella and synedra (12). However many classes of algae, e.g. chlorella, are too small to be removed, even on fine screens (23 microns) and excessive loadings (up to 2×10^6 algae per ml) make this application a limited one.

The parameters of mesh size, submergence, allowable headloss and drum speed [rpm = peripheral speed ÷ (diameter)] are sufficient to determine the flow capacity of a microscreen with given suspended solids characteristics (13).

TABLE 8-5

MICROSCREEN DESIGN PARAMETERS

Item	Typical Value	Remarks
Screen Mesh	20-25 microns	Range 15-60 microns
Submergence	75 percent of height 66 percent of area	
Hydraulic Loading	5-10 gpm/sq ft of submerged drum surface area	
Head-loss (HL) through Screen	3-6 in	Maximum under extreme condition: 12-18 in. Typical designs provide for overflow weirs to bypass part of flow when head exceeds 6-8 in.
Peripheral Drum Speed	15 fpm at 3 in. (HL) 125-150 fpm at 6 in. (HL)	Speed varied to control extreme maximum speed 150 fpm
Typical Diameter of Drum	10 ft	Use of wider drums increases backwash requirements.
Backwash Flow and Pressure	2 percent of throughput at 50 psi 5 percent of throughput at 15 psi	

Among these parameters peripheral speed, hydraulic loading and major variations in mesh size also affect performance on a given feed flow. In addition, drum speed and diameter affect the wastewater flows and pressures needed to effect proper cleaning of the screen.

8.3.3 Hydraulic Capacity

The *filterability index* developed by Boucher (14) quantifies the effect of the feed solids characteristics on the flow capacity of a particular fabric. Boucher assumed that at any constant laminar flow rate the headloss, ΔP in ft, through any given strainer fabric would increase exponentially with the volume passed per unit area (V in cu ft/sq ft):

$$\frac{\Delta P}{\Delta P_o} = e^{IV}$$

In the above relation the filterability index is the exponential rate constant I (in 1/ft).

From the filterability index concept Mixon (13) developed hydraulic capacity relations for continuous operation of a rotating drum microscreen, which can be expressed as follows:

$$\mu = \frac{Q}{A} = \frac{\ln\left[\left(\frac{\Delta P}{C_F}\right)\left(\frac{I\varphi}{R}\right) + 1\right]}{\left(\frac{I\varphi}{R}\right)}$$

Where:

μ = mean flow velocity through submerged screen area (fps)
Q = total flow through microscreen (cfs)
A = submerged screen area (sq ft)
P = pressure drop across screen (ft)
C_F = fabric resistance coefficient (ft/fps or sec) (clean fabric headloss at 1 fps approach velocity)
I = filterability index (1/ft)
φ = decimal fraction of screen area submerged
R = drum rotational speed (rpm)

The expression $\Delta P/C_F$ represents the initial flow velocity through the clean screen as it enters submergence. C_F is a particular characteristic of the screen fabric, varying inversely with mesh opening size as follows:

Mesh Size mu	Fabric Resistance, C_F ft/fps
15	3.6
23	1.8
35	1.0
60	0.8

Limits on ΔP reflect screen fabric mechanical strength and expected operating conditions for the unit. A typical value is 0.5 ft at normally-expected maximum flow.

The relation of parameters in the expression ($I\phi/R$) shows that the effect of a higher index or faster buildup of headloss on the screen may be offset by maintaining a higher drum rotational speed.

Figure 8-8 is a graphical representation of the above relation which Mixon obtained by plotting Q/A against $\Delta P/C_F$ for various values of the parameter $I\phi/R$.

The graph shows lines of constant value for the ratio

$$E = \frac{Q/A}{\Delta P/C_F}$$

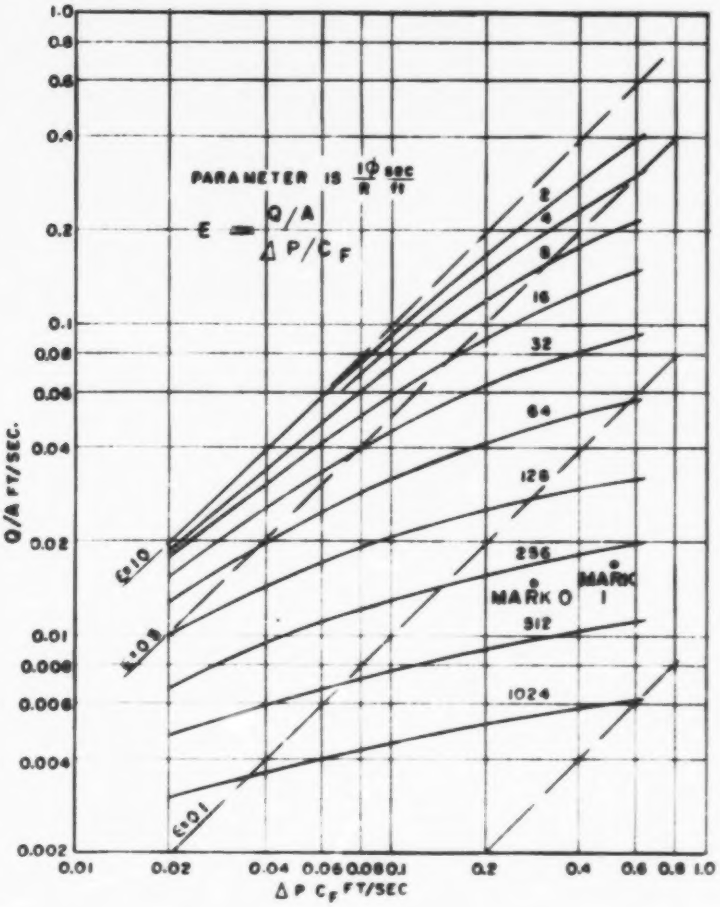
which is the ratio of the mean velocity through the screen to the initial velocity when the screen enters submergence. Recognizing the effect of drum speed on performance Mixon suggested selecting $I\phi/R$ to keep the ratio E below 0.5. Above this limit he assumes that insufficient opportunity is given for a mat to form on the drum and solids removal efficiency is likely to suffer.

The filterability index may be determined by Boucher's laboratory procedure (14), by field testing with an apparatus available from Crane Co. (11) or by analysis of test data from pilot microscreen units using relations such as those proposed by Mixon or Boucher. In some cases where numerous values of 'I' have been obtained, a relationship to influent SS loading may be obtained (15). Solids loading limitations such as those noted by Lynam, et al, (16) would have broader applicability if they were related to a maximum filterability index under which a particular microscreen could maintain a given capacity.

8.3.4 Performance

Suitable relationships have not been developed for quantitative predictions of microscreen performance from knowledge of influent characteristics and key design parameters. Where performance must be predicted closely, pilot studies should be made. Where close prediction is less critical, performance data from other locations with generally similar conditions may serve as a guide.

Table 8-6 provides performance data for a number of microscreen installations for SS removal from secondary effluents including the first such installation made at Luton Sewage Works, England, in the early 1950's. Table 8-7 lists additional American installations provided by two manufacturers.



TYPICAL DESIGN RANGE

5 - 10 GPM / SQ FT

FABRIC	MESH - MU
MARK O	23
MARK I	35

FIGURE 8-8

MICROSCREEN CAPACITY CHART(13)

..E 8-6

MICROSCREEN INSTALLATIONS

LOCATION	Country	Influent Source	Drum Dia x Width ft	Drum Screen Mesh mu	No. of Units	Hydraulic load on Submerged Area Max. gpm/sq ft	Hydraulic load on Submerged Area Avg. gpm/sq ft	Plant Flow Max. mgd	Plant Flow Avg. mgd	Average Suspended Solids Influent mg/l	Average Suspended Solids Effluent mg/l	Removed percent	Reference
Luton	England	Trickling Filter	------,	60	-	9.0	-	5	-	14	8	45	19
Bracknell	England	Trickling Filter	7.5x5,	35	2	6.3	2.2	6.3	2.2	20 16.2	11 6.9	45 57	19
Hambledon R.D.C. Elmbridge	England	Trickling Filter	------, ------,	35 23	- -	10.8 10.8	2.5 2.5	2.5 2.5	0.6 0.6	14 14	8 7.7	43 45	19
Leighton-Linslade U.D.C.	England	Trickling Filter	------,	35	-	6.8	3.9	2.0	1.2	29	11	60	19
Fleet U.D.C.	England	Trickling Filter	10x10,	35	2	6.0	2.0	3.6	1.2	15	6	60	9,19
Esher U.D.C.	England	Trickling Filter	10x10,	-	3	-		10.8 (Design Flow)		19	9	60	9
Hatfield R.D.C.	England	Trickling Filter	7.5x5,	-	3	-		2.55 (Design Flow) 9.0 (Design Flow)		14	8	43	9,19
The Borough of Bury St. Edmunds	England	Trickling Filter	10x10,	23	5	-	-	5.3 (Max. Flow to Date)		28	7	75	9
Franklin Township STP Murraysville, Pa.	U.S.A.	Trickling Filter Final-Settling Tanks	10x10,	23	2	7.6	-	4.0	-	37	6	83+10	9
Letchworth	England	Act.Sludge Final Clarifiers	5x3,	23	1	4.3	3.3	Pilot Study		17	6.6	62	9,19,20
Basingstoke	England	Act.Sludge Final Clarifiers	10x10,	23	5	1.5	1.0	3.2	2.2	13.1	3.9	70	9,19
Euclid, Ohio	U.S.A.	Act.Sludge with Chem.Precip.in Primary Clarif (FeCl)	2.5x2,	23	1	2.5	1.25	Pilot Study 40 gpm	20 gpm	54	8	85	17
Euclid, Ohio	U.S.A.	Act.Sludge with Chem.Precip.in Final Clarif. (FeCl)	2.5x2,	23	1	2.5	1.25	Pilot Study 40 gpm	20 gpm	38	10	74	17
Euclid, Ohio	U.S.A.	UNOX Final Clarifiers	2.5x2,	23	1	2.5	1.25	Pilot Study 40 gpm	20 gpm	65	21	68	17
Lebanon, Ohio	U.S.A.	Act.Sludge Final Clarif.	5x1, 5x1,	35 23	1 1	- -	- -	-	1	27 17	7 2	73 83	15 15
Hanover Park, Ill.	U.S.A.	Act.Sludge Final Clarif.	10x10,	23	1	5.3	2.6	1.5	0.8	6-28	4-11	55(ave.)	16
MSD North Side STP Chicago, Ill	U.S.A.	Act. Sludge Final Clarif.	12.5x30,	23		-		15 (Design Flow)		10	3	67	9

TABLE 8-7

MUNICIPAL MICROSCREENER INSTALLATIONS

Courtesy Zurn Industries

Location	Plant Flow mgd	Units No.	Unit Sizes D x L, ft	Straining Media* microns
Arthur Bloom Apts. Lancaster, Pa.	0.1	2	4x2	20 (polyester)
Ecological Utilities North Miami, Fla.	2.7	1	10x10	20 (polyester)
City of Murfreesboro Murfreesboro, Tenn.	4.0	2	10x10	20 (polyester)
MSD Chicago Lemont, Ill.	--	2	10x10	21
City of Cookeville Cookeville, Tenn.	7.2	2	10x10	21
City of Dayton Dayton, Tenn.	5.4	2	10x10	20 (polyester)
Department of Public Works Erie, Pa.	45.0	3	10x15	21
Village of Pepperpike Pepperpike, Ohio	1.0	1	6x6	21
Borough of Bellefonte Bellefonte, Pa.	3.0	3	6x6	21
Good Samaritan Hospital Islip, New York	0.2	1	4x4	21

* All fabrics stainless steel unless otherwise indicated

TABLE 8-7 (continued)
MUNICIPAL MICROSCREENER INSTALLATIONS

Location	Plant Flow mgd	Units No	Unit Sizes DxL, ft	Straining Media* microns
Borough of Carroltown Carroltown, Pa.	1.0	2	4x4	21
Cincinnati Dept. of Sewers Cincinnati, Ohio	0.5	1	4x4	21
Muncie Mall Muncie, Indiana	0.2	1	4x4	21
Opalaka Sewage Plant Chesterland, Ohio	0.3	1	4x4	21
I.B.M. Essex Junction, Vt.	0.2	1	4x4	35
Hot Springs Village Hot Springs, Arkansas	0.3	1	4x4	21
	(Courtesy of Crane Co.)			
Union "76" Oil Co. Clarion, Pa.	0.1	1	5x1	23
Deer Creek State Park, Ohio	0.3	1	5x3	23
Oakbourne Hospital, Westchester, Pa.	0.1	1	5x1	23
Sugar Creek STP, Greene County, Ohio	6.0	2	10x10	23
Little Miami STP, Greene County, Ohio	6.0	2	10x10	23
Westminster, Md.	3.0	1	10x10	23
Hammond, Ind.	1.5	1	7.5x5	23
East Wheatfield Township, Pa.	0.1	1	5x1	23

TABLE 8-7 (Cont'd)
MUNICIPAL MICROSCREENER INSTALLATIONS

Location	Plant Flow mgd	Units No.	Unit Sizes D x L, ft.	Straining Media* microns
Hartville, Ohio	2.5	2	7.5x5	23
Louisville, Ky.	0.8	2	5x3	60
Browntown, Minn.	0.1	1	5x1	23
Salisbury, N.C.	6.0	2	10x10	23
Allen County, Ohio	2.5	2	7.5x5	23
Fairmont, Minn.	6.0	2	10x10	23
Parkway, Md.	15.0	5	10x10	23
Penn State University, Pa.	0.3	1	5x3	23
Bolingbrook, Illinois	0.5	1	7.5x5	60
Park Forest, Illinois	2.0	2	7.5x5	23
Sugar Creek, Ohio	0.3	1	5x3	23
Union Oil Co. Harrisburg, Pa.	0.1	1	5x1	23
Chicago Metro Sanitary District Chicago, Illinois	2.0	1	10x10	23
Ursuline Academy, Ohio	0.3	1	5x3	35
Jackson Township, N.J.	0.1	1	5x1	35
FWPCA Research Project, Phila.	0.2	1	5x3	23
Akron STP, Ohio	3.0	1	7.5x5	60
Wm. Henry Apts., Dowington, Pa.	0.2	1	5x1	35

TABLE 8-7 (continued)

Location	Plant Flow mgd	Units No.	Unit Sizes D x L, ft	Straining Media microns
Franklin Twp., Pa.	4.0	2	10x10	23
Hempfield Twp., Pa.	6.0	2	10x10	35
Chelsea Ridge Apts., N.Y.	0.2	1	5x3	23
Harpeth Valley, Tenn.	1.5	1	7.5x5	23
University School Cleveland, Ohio	0.1	1	5x1	23
Willoughby Hills Cleveland, Ohio	0.1	1	5x1	23
Lionville, Pa.	0.7	2	5x3	23
Margate STP, Fla.	3.5	2	10x10	23
Lauderhill STP, Fla.	3.5	2	10x10	23
Bel-Aire STP. Miami, Florida	0.9	1	7.5x5	23
Homestead, Fla.	0.2	2	5x1	23
Upper Sandusky, Ohio	1.5	2	7.5x5	23
Petosky, Michigan	2.5	2	10x10	23
Ravenna, Ohio	1.5	2	7.5x5	23
Bolingbrook STP, Ill.	0.5	1	7.5x5	60
Commonwealth Edison, Ill.	2.0	1	10x10	23
Lucas County, Ohio	0.2	2	5x1	23
Pymatuning State Park, Pa.	0.1	1	5x1	35

Figure 8-9 shows average operating results from a number of British tertiary microscreen installations with various hydraulic loadings.

Figure 8-10 presents the results of three extended British studies on microscreening of trickling filter and activated sludge secondary effluents.

Some general conclusions can be made about the microscreen as a device for removing SS from secondary effluents:

1. Under best operating conditions microscreen units can reduce solids to as low as 5 mg/l.

2. Although the SS removal pattern is irregular, performance tends to be better at lower hydraulic loadings (Figure 8-10a).

3. Increases in influent suspended solids are reflected in the effluent but with noticeable damping of peaks (Figure 8-10).

4. Microscreens are applicable in place of clarifiers to polish effluent from low rate trickling filters, if the solids are generally low in concentration and well flocculated (Figure 8-10b).

Other investigations provide insights beyond those cited above:

Data from Lebanon, Ohio, (15) show better removal with smaller mesh sizes.

Commenting on how solids characteristics affect microscreen performance, one British study (8) notes that a clear non colloidal secondary effluent containing a reasonable amount of suspended solids would result in a better effluent than colloidal effluent containing less suspended solids.

Chemical application can unfavorably alter solids characteristics: Lynam, et al. (16) reported poor removals when applying an alum flocculated secondary effluent directly to a microscreen. In contrast, at Euclid, Ohio (17) a microscreen removed 74 to 85 percent of the solids from the settled effluent of an activated sludge pilot plant with mineral addition for phosphorus removal. The better performance at Euclid could be attributed to a tougher biological nature of the effluent solids.

Microscreen suppliers (9) (18) stress the importance of minimizing shearing action on microscreen influent to avoid breaking up flocculated particles. This is advanced as a reason for settling limits on drum peripheral speed and for avoiding pumping ahead of microscreen units.

Lynam, et. al, (16) indicate that microscreening at lower drum speeds yields better quality effluent. This is attributed to better straining action through the thicker mat of solids which builds up at low speeds.

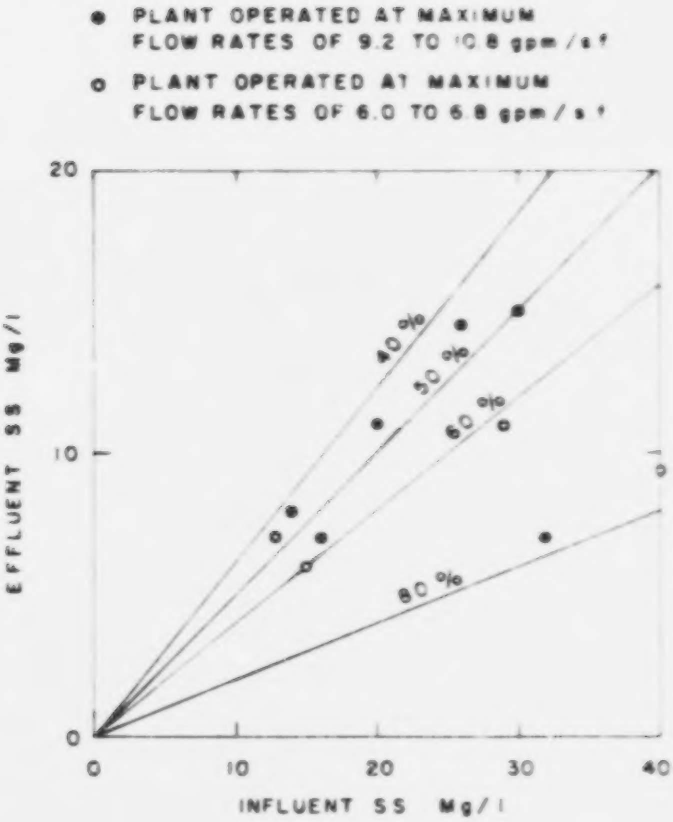
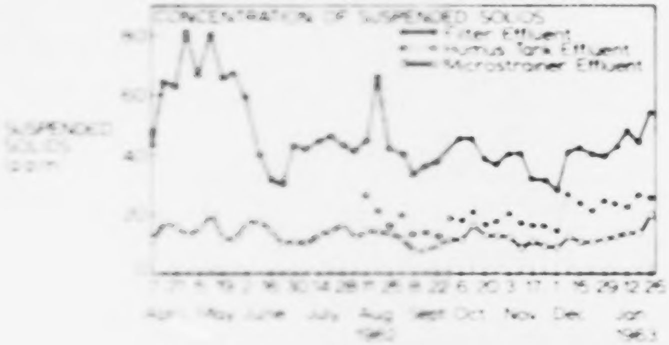


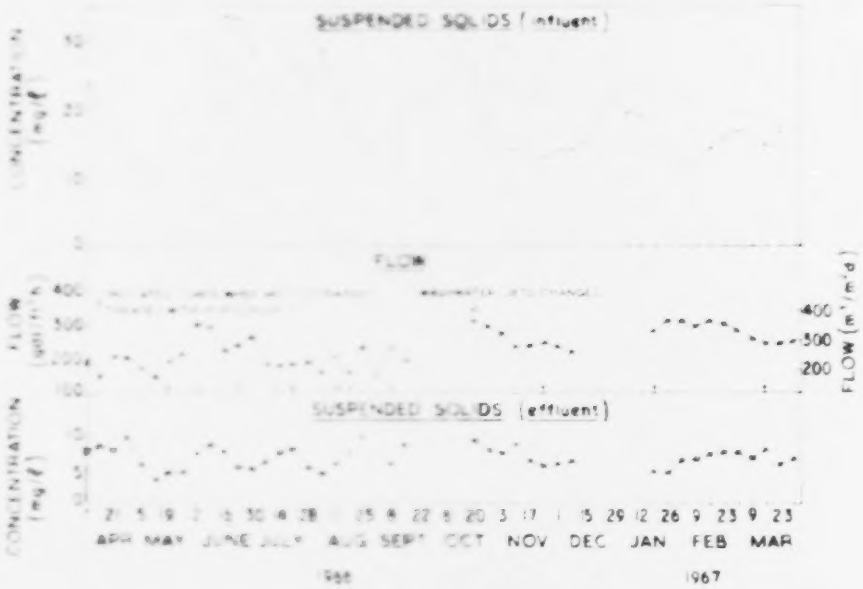
FIGURE 8-9

MICROSCREEN REMOVAL AT VARIOUS FLOW RATES
(After Isaac and Hibbert (19))



A. HARPENDEN SEWAGE WORKS

(Courtesy of Crane Co.)

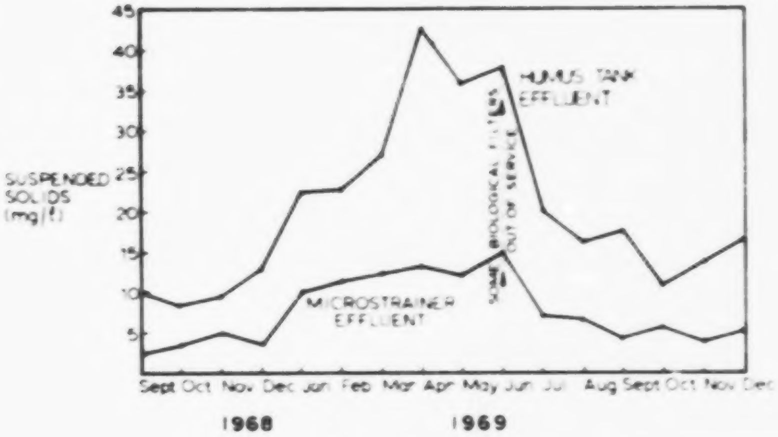


B. LETCHWORTH WATER POLLUTION CONTROL WORKS

Data were obtained from analyses of continuous records, each point representing weekly mean hourly readings. Broken lines indicate break in operation of plant

FIGURE 8-10

MICROSCREENING OF TRICKLING FILTER PLANT EFFLUENT



C. FLEET U.D.C.
(Courtesy of Crane Co.)

FIGURE 8-10(Continued)

As shown in Figure 8-8 under peripheral speed and hydraulic loading limits set by manufacturers and regulatory authorities, the ratio E for most microscreen designs actually falls below 0.1, a lower limit suggested by Mixon for most efficient utilization of equipment.

The wide range of suggested design values underlines the need for developing quantitative relations between removal efficiency and key design parameters.

8.3.5 Microscreen Construction

The basic screen support structure is a drum shaped, suitably stiffened rigid frame supported on bearings to allow rotation. Designs using water lubricated axial bearings or greased bearings located on the upper inside surface of the rotating drum allow submergence well above the central axis.

Both plastic (polyester) and stainless steel are used for the microscreen media itself. Greater mechanical strength, especially at higher temperatures, is the prime advantage stated for stainless steel (9) (18). Greater economy and chemical resistance are pointed to as advantages of plastic (18).

Depending on manufacturer, screen fabric is supplied either in small sections (8 in. x 12 in.) supported by and fastened directly to the drum frame or in larger (18 in. x 24 in.) panels consisting of fabric integrally bonded to a grid like supporting mesh of stainless steel. These panels are attached to the drum frame.

One manufacturer offers a microscreen unit with an accordion or pleated outer surface to achieve up to 30 percent more filtering surface within the same general dimensions of regular designs. The unit is 12.5 feet by 30 feet and has a rated capacity of 15 mgd. This unit is shown in Figure 8-11.

In the past cast bronze and cast carbon steel were used as drum and frame construction materials. The present practice is to use fabricated carbon steel. Generally, smaller units are factory assembled in steel tanks while large units are placed in concrete structures.

Table 8-8 illustrates the approximate size and power requirements for various microscreen units.

8.3.6 Screen Fabric and Operating Headloss

Microscreen fabrics normally are woven of stainless steel or plastic (polyester with polypropylene supporting grid) with openings in the range of 15 to 60 microns.

Plastic fabric is less subject to chemical attack by strong chlorine or acid cleaning solutions. Stainless steel can better withstand temperatures encountered in steam cleaning.

Suggested operating headloss limits for microscreens are based on observation of the effect of differential head on screen life. Standard design calls for a 3 in. headloss at average flows

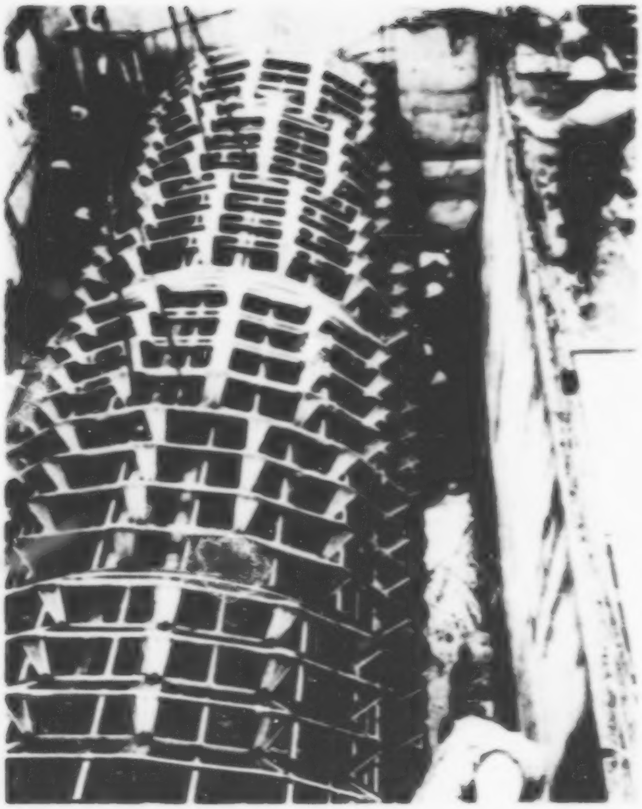


FIGURE 8-11

MICROSCREEN UNIT WITH PLEATED OUTER SURFACE
(Courtesy of Cochrane Div., Crane Co.)

TABLE 8-8

TYPICAL MICROSCREEN POWER AND SPACE REQUIREMENTS

	Drum Sizes		Floor Space		Motors		Approx. Ranges
Source	Diam.	Length	Width	Length	Drive	Wash Pump	of Capacity
Code	ft	ft	ft	ft	BHP	BHP	mgd
A	5.0	1.0	8	6	0.50	1.0	0.07 - 0.15
A	5.0	3.0	9	14	0.75	1.0	0.2 - 0.4
A	7.5	5.0	11	16	2.00	5.0	0.5 - 1.0
A	10.0	10.0	14	22	5.00	7.5	1.5 - 3.0
B	4.0	4.0	[illegible]	15	0.75	1.0	0.2 - 0.4
B	6.0	6.0	10	17	2.00	1.5	0.5 - 1.0
B	10.0	10.0	14	22	5.00	5.0	1.5 - 3.0

Code A - Courtesy Crane Co., Cochrane Div.
Code B - Courtesy Zurn Industries

and 6 in. for normally-expected maximum flows (9) (18). For occasional peaks (less than 3 percent of time) headlosses up to 24 in. can be tolerated. Crane Co. indicates that stainless steel screens operated under the above conditions would have a life of 10 years; if operated continuously at a 24 in. headloss the same screen might only last 6 months. (9)

8.3.7 Hydraulic Control

Hydraulic control of microscreening units is effected by varying drum speed in proportion to the differential head across the screen. The controller is commonly set to give a peripheral drum speed of 15 fpm at 3 in. differential and 125 to 150 fpm at 6 inches (9) (18). In addition, backwash flow rate and pressure may be increased when the differential reaches a given level (9) (18).

The operating drum submergence is related to the effluent water level and headloss through the fabric. The minimum drum submergence value for a given installation is the level of liquid inside the drum when there is no flow over the effluent weir. The maximum drum submergence is fixed by a bypass weir which permits flows in excess of unit capacity to be bypassed; at maximum submergence the maximum drum differential should never exceed 15 inches.

Effluent and bypass weirs should be designed as follows:

1. Select drum submergence level (70 to 75 percent of drum diameter) for no flow over the effluent weir.

2. Locate top of effluent weir at selected submergence level.

3. Determine maximum flow rate.

4. Size effluent weir to limit liquid depth in effluent chamber above the weir to 3 in. at the maximum flow rate.

5. Position the bypass weir 9 to 11 in. above effluent weir. (3 in. head on effluent weir maximum flow plus 6 to 8 in. differential on drum at maximum drum speed and maximum flow).

6. Size bypass weir length to prevent the level above effluent weir flow exceeding 12 to 18 in. at peak maximum flow or overflowing the top of the backwash collection hopper.

8.3.8 Backwashing

Backwash jets are directed against the outside of the microscreen drum as it passes the highest point in its rotation. About half the flow penetrates the fabric,dislodging the mat of solids formed on the inside (15). A hopper inside the drum receives the flushed-off solids. The hopper is positioned to compensate for the trajectory that the solids follow at normal drum peripheral velocities.

Microscreen effluent is usually used for backwashing. Straining is required to avoid clogging of backwash nozzles. The in-line strainers used for this purpose will require periodic cleaning; the frequency of cleaning will be determined by the quality of the backwash water.

The backwash system used by Zurn employs two header pipes, one operates continuously at 20 psi, while the other operates at 40 psi when the unit receives a high solids loading. The Crane system also uses two sets of jets but both operate continuously at pressures from 15 to 55 psi. Under normal operating conditions these jets operate at 35 psi. Once a day they are operated at 50 psi for 1/2 hour to keep the jets free of slime build-up. Should this procedure fail to keep the jets clean, the pressure is raised to 55 psi. At this pressure the spring loaded jet mouth widens to allow for more effective cleaning.

Backwash pressure is also increased to compensate for heavy solids loadings which require the higher pressure for thorough cleaning. Crane reports that no major problems have been encountered with this jet design (9).

Prior to 1967 Crane designed backwash systems to operate only at 15 psi. A pilot study in Letchworth England (20) showed the superiority of the higher pressure system. Results of this study showed:

1. Operation at 50 psi, as opposed to 15 psi, increased the process flow capacity 30 percent.

2. Suspended solids concentration in the backwash increased from 260 mg/l at 15 psi to 425 mg/l at 50 psi.

3. Water consumption of the jets as a percent of process effluent decreased from 5 percent at 15 psi to 2 percent at 50 psi.

In general, backwash systems are operated at as low a pressure as possible consistent with successful cleaning. High pressure operation incurs added system maintenance, particularly jet replacement, and is used only as needed.

8.3.9 Supplemental Cleaning

Over a period of time screen fabrics may become clogged with algal and slime growths, oil, and grease. To prevent clogging, cleaning methods in addition to backwashing are necessary.

To reduce clogging from algae and slime growth, Crane Company recommends the use of ultraviolet lamps placed in close proximity to the screening fabric and monthly removal of units from service to permit screen cleaning with a mild chlorine solution. While most literature sources say ultraviolet lamps are of value, one authority (21) feels these lamps are uneconomical because they require frequent replacement. Zurn Industries claims that, because their screening fabric is completely bonded to the supporting material, crevices where algae become lodged are eliminated and backwashing alone is sufficient to remove algal and associated slime growths (18).

Where oil and grease are present, hot water and/or steam treatment can be used to remove these materials from the microscreens. Plastic screens with grease problems are cleaned monthly (9) with hot water at 120° F (18) to prevent damage to the screen material. Down time for cleaning may be up to 8 hours.

8.3.10 Operation

In starting a microscreening unit care should be taken to limit differential water levels across the fabric to normal design ranges of 2 to 3 inches. For example, while the drum is being filled it should be kept rotating and the backwash water should be turned on as soon as possible. This is done to limit the formation of excessive differential heads across the screen which would stress the fabric during tank fill up.

Leaving the drum standing in dirty water should be avoided because suspended matter on the inside screen face which is above the water level may dry and prove difficult to remove. Low flow bypass (introducing unscreened waters such as plant overloads) into the microscreen effluent compartment should also be avoided (18).

If the unit is to be left standing for any length of time the tank should be drained and the fabric cleaned to prevent clogging from drying solids.

8.4 Other Screening Devices

Conventional static screens have not been used with success in municipal wastewater treatment. Recently, however, a centrifugal screen, the Sweco Concentrator, has demonstrated its effectiveness. In this unit, influent is directed against the inside of a rotating cylindrical screen cage (See Fig. 8-12). It is claimed that the rotational speed (centrifugal force of 3 to 6 gravities) increases hydraulic capacity and, together with the impingement angle, permits separation of solids finer than the screen openings (150 to 105).

Separated solids and the rejected portion of the liquid flow are removed from the bottom of the unit while effluent is taken off at the periphery. Screen blinding is cleared by timer-actuated spray cleaning systems which direct water jets against both the inner and outer screen surfaces.

At Contra Costa County, California, a 60 inch unit treats 0.9 mgd wastewater containing about 200 mg/l SS (22). The waste flow is split into two streams, a small volume (15 percent) concentrated stream and a supernatant (85 percent of influent) stream. The concentrated stream is settled in the conventional primary basin (formerly overloaded but now capable of good solids reduction at the lower hydraulic loading). The supernatant is treated by flotation (the Sweco concentrator does not specifically remove floatables but thoroughly aerates the wastewater aiding subsequent flotation) and finally, flotation and settled concentrate effluents are mixed and chlorinated before disposal. Overall SS reductions for the system including concentration, flotation, settling and chlorination are reported as 70 to 80 percent. Similar removals are claimed for settled secondary effluent from aeration processes (23). Design flow rates are claimed to range from 40 to 100 gpm/sq ft. The con-

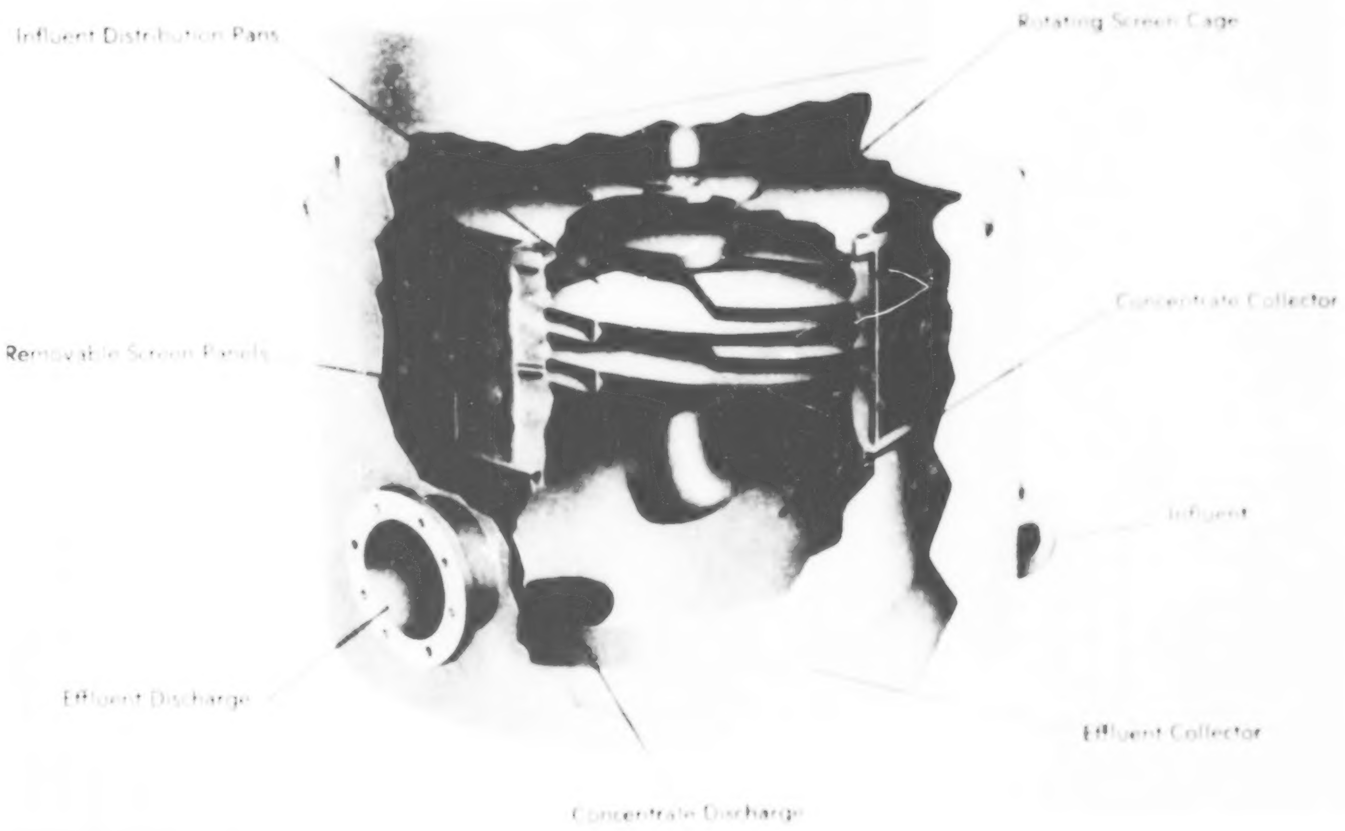

Influent Distribution Pans
Rotating Screen Cage
Concentrate Collector
Removable Screen Panels
Influent
Effluent Discharge
Effluent Collector
Concentrate Discharge

FIGURE 8-12

THE SWECO CONCENTRATOR
(Courtesy of SWECO, Inc.)

centrator is of epoxy-lined steel and screen construction is of stainless steel or polyester in plastic frames. During operation the outside of the screen is washed every 20 minutes with cold clarified water and the inside with 5 to 7 gpm hot water.

8.5 Diatomaceous Earth Filters

Diatomaceous earth (DE) filters have been applied to the clarification of secondary effluents at pilot scale. No full-scale installations have been characterized in the literature. They produce a high quality effluent but appear unable to handle the solids loadings normally expected in this application.

DE filtration utilizes a thin layer of precoat formed around a porous septum to strain out the suspended solids in the feedwater which passes through the filter cake and septum. The driving force can be imposed by vacuum from the product side or pressure from the feed side. As filtration proceeds, headloss through the cake increases due to solids deposition until a maximum is reached. The cake and associated solids are then removed by flow reversal and the process is repeated. In the cases where secondary effluents have been treated by this process, a considerable amount of diatomaceous earth (body feed) has been required for continuous feeding with the influent in order to prevent rapid buildup of headlosses. Generally, the DE filtration process is capable of excellent removal of suspended solids but not colloidal matter.

A wide variety of diatomaceous earth (diatomite) grades are available for use. As might be expected, the coarser grades have greater permeability and solids-holding capacities than do the finer grades which will generally produce a better effluent. Some grades of diatomite are pretreated to change their characteristics for improved performance. A number of vessel configurations are available, with open-basin vacuum and vertical pressure designs most common. (See Figures 8-13, and 8-14.)

Design criteria for diatomite filters have been discussed by Bell (24). The filtration cycle can be divided into two phases, run time and down time. Down time includes the periods when the dirty cake is dislodged from the septum and removed from the filter and when the new precoat is formed. Run time commences when the feed is introduced to the filter and ends when a limiting headloss is reached. The single most important factor in secondary effluent filtration by DE filters is the amount of body feed required during the filtration or run time. The body feed rate is the largest operating cost factor and strongly affects the operating economics of the process. Similarly, it is related to cycle time between backwashing which determines the installed filtering area, hence the capital cost economics. A ratio of 5 to 6 mg/l of body feed per JTU of influent turbidity was required at San Antonio and the possible need for a higher ratio was suggested (25). Another pilot study used a variety of ratios and filtration rates and used both pressure and vacuum systems for secondary effluent filtration (26). Some results from this study are shown in Table 8-9. Both studies indicate that a precoat of about 0.1 lb/sq ft of filter area and greater than 6 mg/l of body feed per JTU of turbidity should be used.

An English study with a reversible-flow DE unit also resulted in uneconomical operating conditions due to excessive body feed, short filter runs and high backwash water

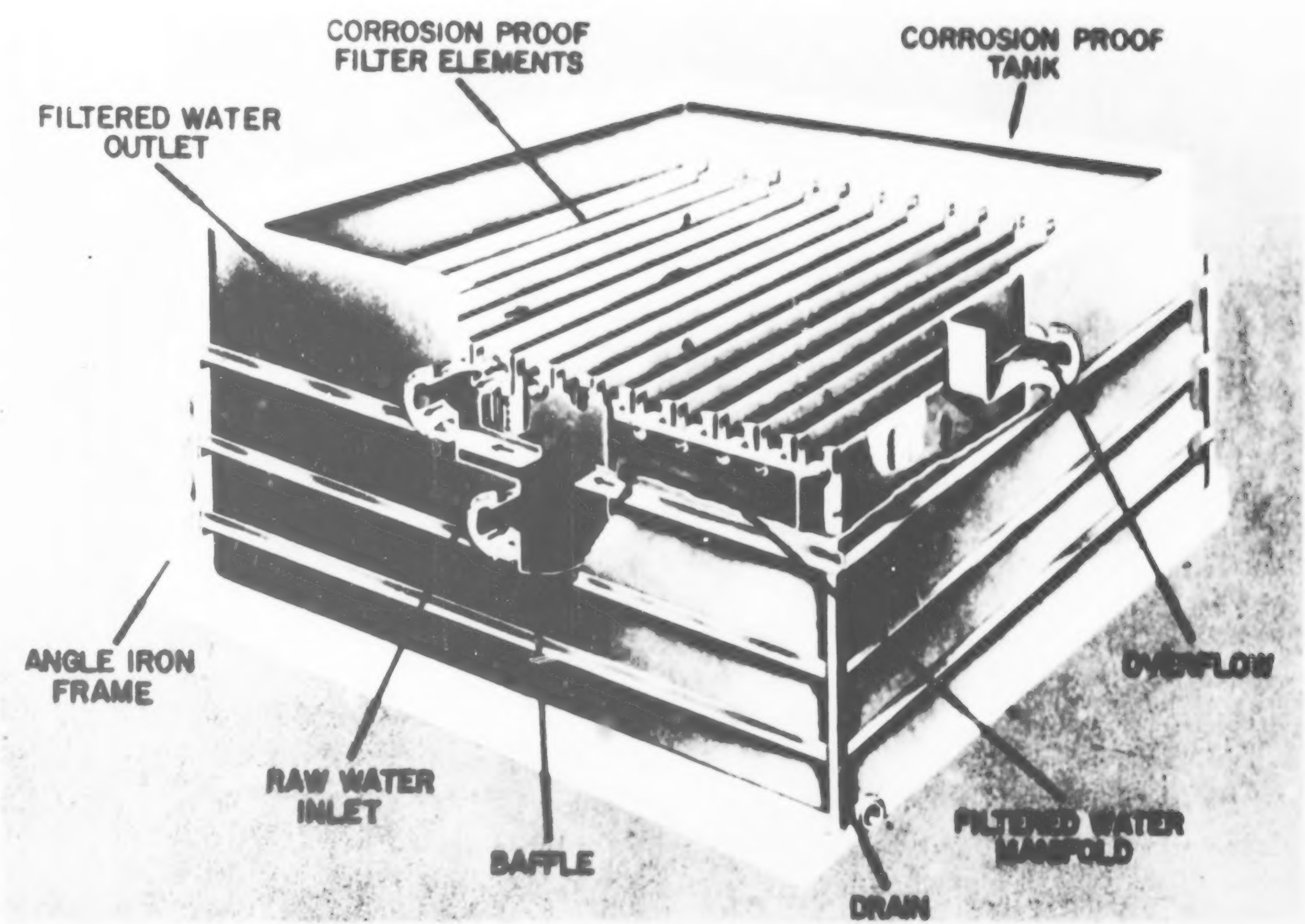

CORROSION PROOF FILTER ELEMENTS
CORROSION PROOF TANK
FILTERED WATER OUTLET
ANGLE IRON FRAME
RAW WATER INLET
BAFFLE
DRAIN
FILTERED WATER MANIFOLD
OVERFLOW

FIGURE 8-13
VERTICAL LEAF VACUUM FILTER
(Courtesy of Johns-Manville)

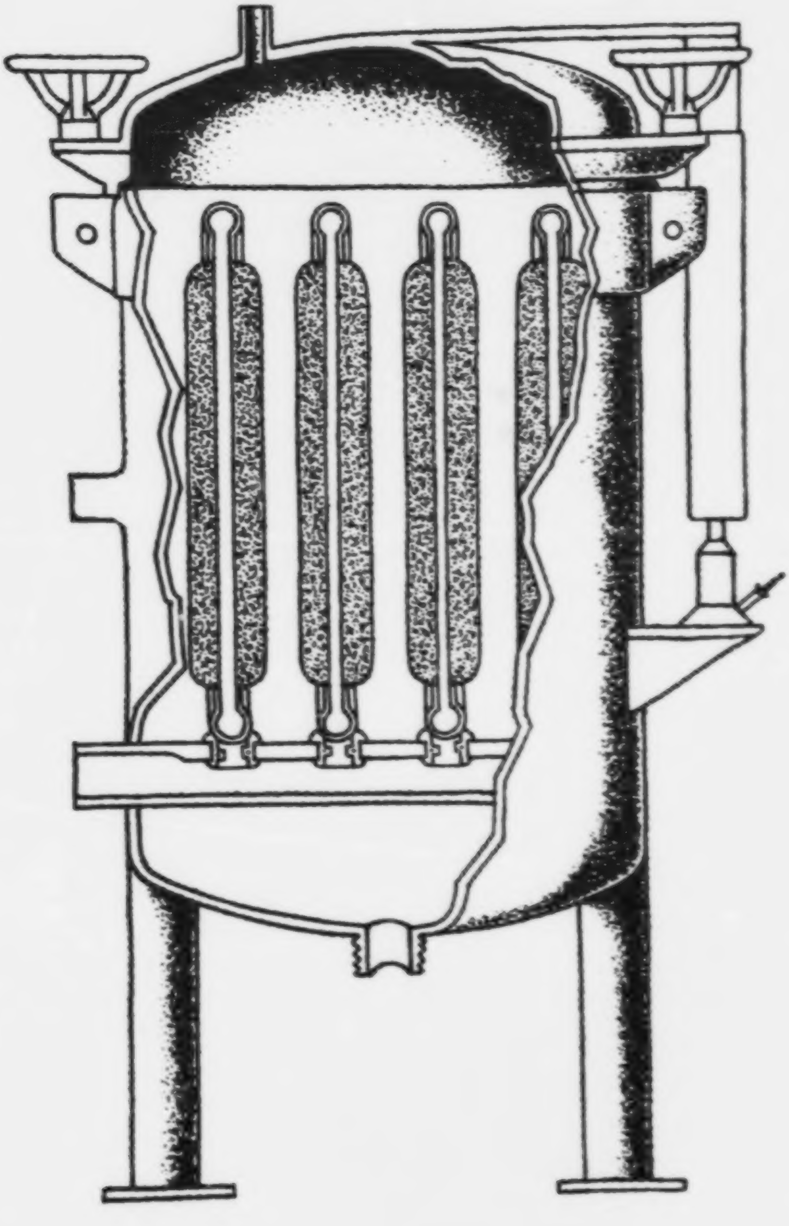


FIGURE 8-14
VERTICAL LEAF PRESSURE FILTER, VERTICAL TANK
(Courtesy of Johns-Manville)

TABLE 8-9

DIATOMACEOUS EARTH FILTRATION OF SECONDARY EFFLUENT

Flow Rate	Body Feed	Turbidity		Run Length	Type
		In	Out		
gpm/sq ft	mg/l	JTU	JTU	hr	
0.53	42	5.5	0.8	19.5	Vacuum
0.75	33	5.2	0.8	10.7	Vacuum
1.0	19	4.4	0.4	5.4	Vacuum
0.50	50	8.2	3.1	50.0	Pressure
0.81	42	8.3	3.9	28.4	Pressure
1.0	45	7.5	3.0	31.0	Pressure

requirements (27). Acceptable operation was possibly only with very low influent solids (3 to 13 mg/l). None of these studies considered the possibility of recovering DE filter aid, which could reduce the estimated costs significantly (28).

8.6 Ultrafiltration

8.6.1 General

Ultrafiltration (UF) is the title given to a form of membrane separation which employs relatively coarse membrane separation at relatively low pressures. The process should be differentiated from reverse osmosis which is a similar process used for dissolved solids separation using fine membranes and high pressures. Ultrafiltration, using a thin semi-permeable polymeric membrane is reported most successful in separating suspended solids as well as large-molecule colloidal solids (0.002 to 10.0 μ) from wastewater (29). Fluid transport and solids retention are achieved by regulating pore size openings. Thus, the ultrafiltration process is a physical screening through molecular-sized openings rather than one controlled by molecular diffusion.

A system employing high-MLSS aeration followed by UF operated at Pikes Peak since 1970 has proven capable of removing virtually 100 percent of the suspended matter and 93 to 100 percent of the associated BOD, COD and TOC from aerated mixed liquors (30).

8.6.2 Application

Several installations (29) (30) (31) have proven the ability of the activated sludge-ultrafiltration process to remove all SS and almost all bacteria and BOD. These systems are typified by Figure 8-15. Results of such installations are given in Table 8-10. Because UF installations have all produced zero SS effluents, other parameters are given to illustrate process capability. More detailed data on the Pikes Peak installation is given in

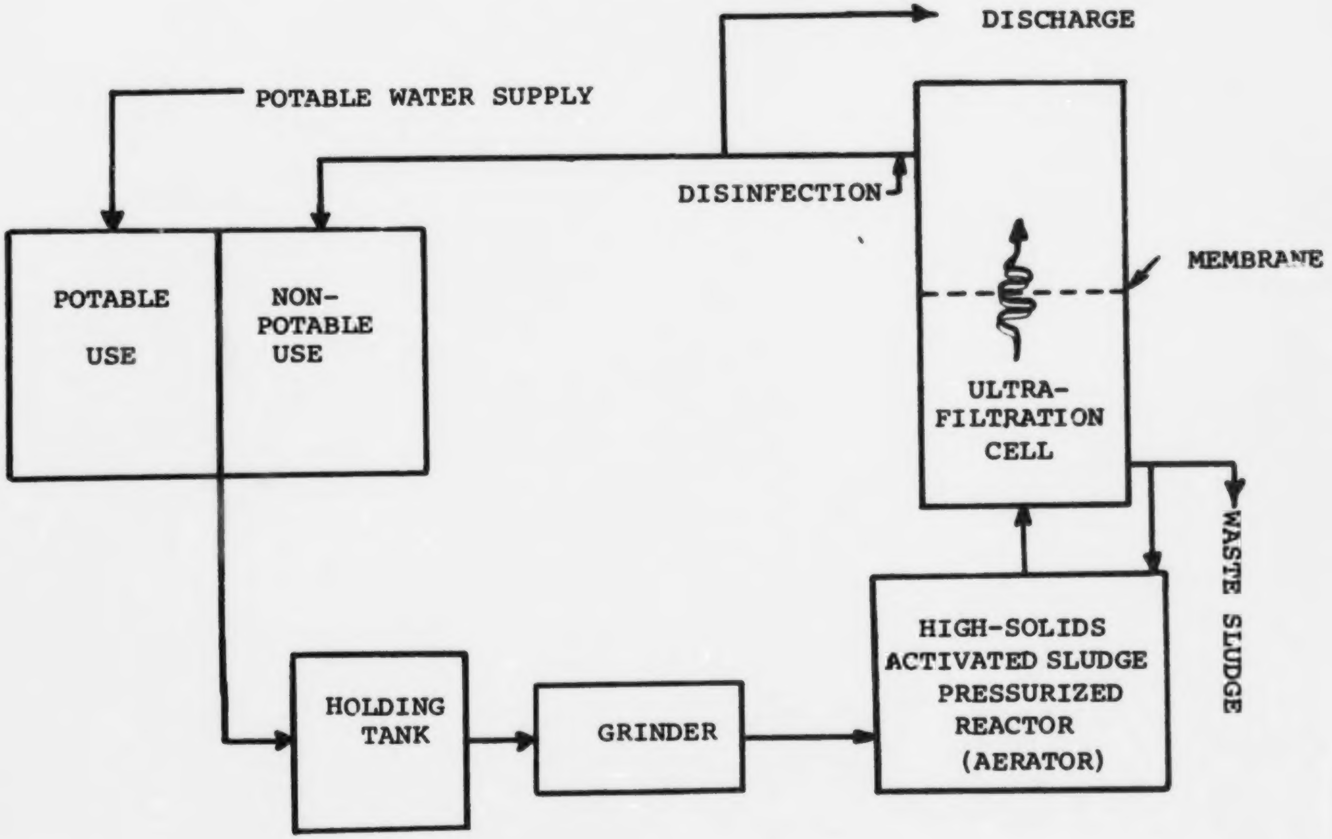

DISCHARGE
POTABLE WATER SUPPLY
DISINFECTION
MEMBRANE
POTABLE USE
NON-POTABLE USE
ULTRA-FILTRATION CELL
WASTE SLUDGE
HOLDING TANK
GRINDER
HIGH-SOLIDS ACTIVATED SLUDGE PRESSURIZED REACTOR (AERATOR)

FIGURE 8-15
SCHEMATIC FLOW DIAGRAM OF THE PIKES PEAK TREATMENT & REUSE SYSTEM

Table 8-11. Coliform counts in all instances were above zero but were attributed to outside contamination rather than passage through the membranes (30) (31).

TABLE 8-10
RESULTS OF ULTRAFILTRATION INSTALLATIONS

	Capacity	Effluent BOD	Coliform	Flux		Pressure	Superficial Flow Rate
	gpd	mg/l	No./100 m	sq ft	final	psi.	fps
Fabric Fire Hose, Sandy Hook	3,600	2-15 (av. 5)	1-100 *	18	8	22-27	4-6
Pikes Peak	21,000	1	0-11 **	30 16	6 9	50	5-6

* includes deliberate upset tests
**attributable to external contamination

The major drawback of ultrafiltration is the high capital and operating costs. Phosphate and color removal are both negligible, but they may not be necessary in many places. The high cost may be offset by compactness where space is a critical factor, such as on a ship or a mountain top. A 6800 gpd shipboard installation was designed to occupy a volume of 6 x 8 x 9 ft (31).

8.6.3 Design

The most important design considerations are:

1. Membrane area
2. Membrane configuration
3. Membrane material
4. Membrane life
5. Driving force.

TABLE 8-11
SUMMARY OF PIKES PEAK DATA

Parameter	CSSTP '70 72 days		Summit '70 22 days		Summit '71 49 days		Summit '72 83 days		Weighted Average 226 days		Percent Removal
	Inf.	Eff.	Inf.	Eff.	Inf.	Eff.	Inf.	Eff.	Inf.	Eff.	
BOD (mg/l)	382	1	285	1	362	1.3	426	6.2	384	3	99.2
COD (mg/l)	678	20	547	32	738	52	839	536	737	40.5	94.6
TOC (mg/l)	192	7.5	136	6.6	197	9.8	–	–	185	8.1	95.6
TSS (mg/l)	323	0	129	0	172	0	263	0	249	0	100
FLUX (gpd/sq ft)	––		30 to 6		16 to 9		––				
MLSS (mg/l)	––		3954		4156		––				
Fecal Coliform (per 100 ml)	–	0	–	0	–	6	–	11		9	

8.6.3.1 Area

Membrane area is a function of flux which is determined by membrane construction and the fouling characteristics of the wastewater. A flux of 8 gpd/sq ft has proven satisfactory at the Pikes Peak installation and can be used as a normal design figure in calculating necessary area. Membrane flux tends to decrease with time due to surface fouling. It has been found that physical elimination of foulants, mostly organic acids and polarized materials, lessens their flux-reducing effect (29). By operating the process at liquid velocities of 3 to 10 fps parallel to the membrane surface, scouring of contaminants can be accomplished and a more stable flux achieved (29). At such velocities, with normal membrane fluxes, single-pass design would require impractically large membrane area. Therefore, the wastewater is recirculated as shown in Figure 8-15. Some blowdown of concentrated waste results to prevent excessive solids buildup. The blowdown can be intermittent, at a rate sufficient to keep the MLSS within acceptable ranges, usually 4,000 to 15,000 mg/l (29) (32).

8.6.3.2 Membrane Configuration

This aspect of design concerns the amount of membrane surface area which can be incorporated into a module. Because of low membrane fluxes it is imperative to design the module to maximize membrane surface area. One configuration adopted solely for ultrafiltration is the storage battery configuration, as shown in Figure 8-16. The membrane is cast on both external faces of a hollow plate. A number of these plates are arranged in a parallel array. The edges of these plates face the incoming stream of solids and act as a coarse screen which can be backwashed by reversing the direction of the approaching flow. Other designs include tubular support elements over which a membrane is wound helically or in which the membrane is enclosed in a continuous spiral.

8.6.3.3 Material

The membrane itself is made up of two basic layers:

1. Surface—an extremely thin homogenous polymer of 0.1 to 10.00 microns (typical, 5.0 microns).

2. Surface support—an open cell of 5 to 10 mil thickness

The membrane, in turn, is supported on a porous sheet (paper) for added mechanical support.

The thin surface layer controls the transport and rejection properties of the membrane. Numerous means and types are available and can be tailored to the particular application. Typical membrane specification ranges are listed in Table 8-12.

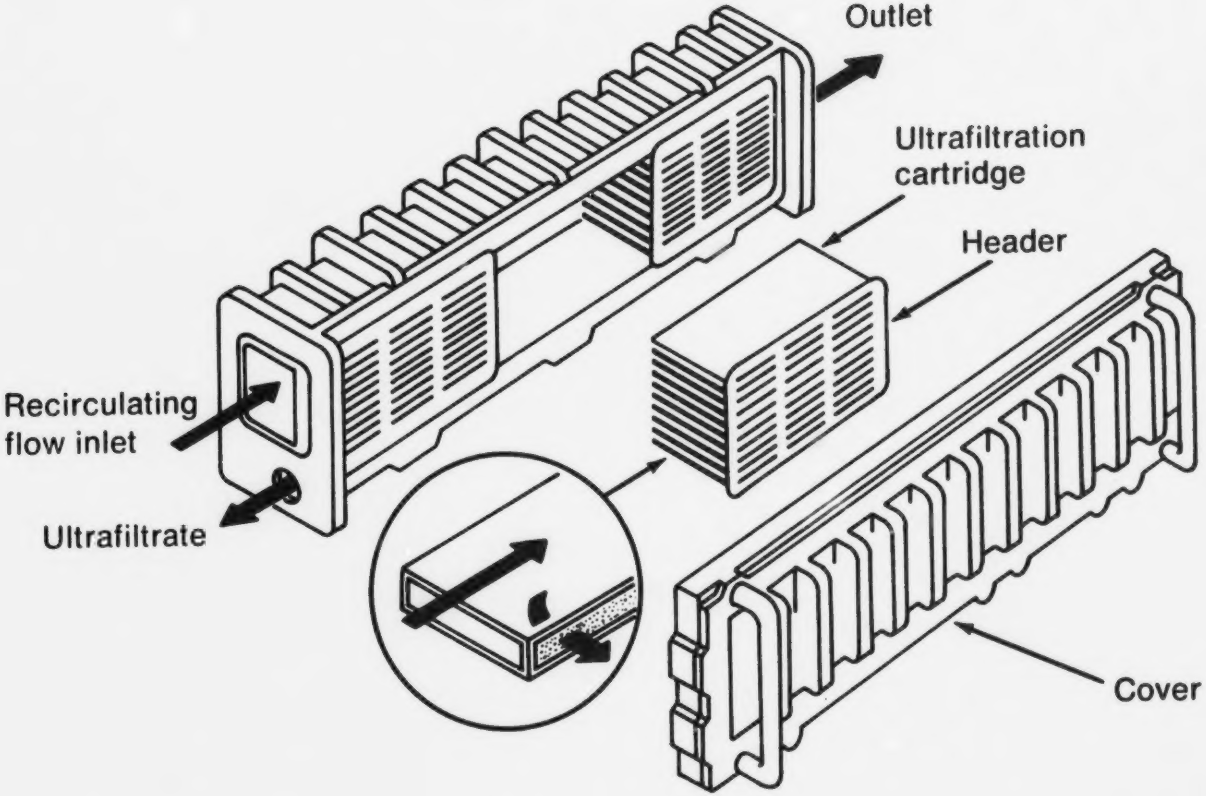

Outlet
Ultrafiltration cartridge
Header
Recirculating flow inlet
Ultrafiltrate
Cover

FIGURE 8-16
"STORAGE BATTERY" MEMBRANE MODULES
(Courtesy of Dorr-Oliver).

TABLE 8-12

TYPICAL MEMBRANE SPECIFICATIONS

Material	Most organic polymers
Water Permeability	7-290 gpd per sq ft at 30 psi
Molecular weight	340-45,000
Retentivity	60-100 percent
Maximum Operating Temperature	50-120ºC

Water permeability is used to characterize the porosity of the membrane, but does not represent the stabilized, long-term flux on a process fluid. In the waste treatment field, fluxes of 7 to 10 gallons per square foot of membrane surface per day are typical (31).

Given the current state of membrane manufacturing technology, almost any set of clean water performance characteristics, without consideration for fouling can be produced. A few of the leading manufacturers of ultrafiltration membranes are Romicon Corp., Abcor, Inc. and Dorr-Oliver, Inc. Catalogues offering a wide variety of membranes are available.

8.6.3.4 Membrane Life

Membrane life is a function of fouling and required flux rates. A membrane may be considered acceptable for a life span of 6 months in continuous operation with an initial flux of 18 and a final flux of 8 gpd sq ft. A plant must be designed for the lower figure and membrane replacement made when the design figure is reached.

8.6.3.5 Driving Force

The driving force for transport of water through the membrane is pressure. Operation is achieved at pressure gradients of approximately 25 psi. Total system pressures do not exceed 50 psi. Very recent work has shown that vacuum extraction of the product can be used advantageously in certain applications (31).

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CHAPTER 9

GRANULAR MEDIA FILTRATION

9.1 Introduction

9.1.1 Applications and General Description

This process, long applied in treatment of municipal and industrial water supplies, is becoming widely used for wastewater treatment both in upgrading existing conventional plants and in designs of new advanced treatment facilities. Next to gravity sedimentation it is the most widely used process for separation of wastewater solids. The following specific applications have been noted (1):

1. Removal of residual biological floc in settled effluents from secondary treatment by trickling filters or activated sludge processes.

2. Removal of residual chemical-biological floc after alum, iron, or lime precipitation of phosphates in secondary settling tanks of biological treatment processes.

3. Removal of solids remaining after the chemical coagulation of wastewaters in tertiary or independent physical-chemical waste treatment.

In these applications filtration may serve both as an intermediate process to prepare wastewater for further treatment (such as carbon adsorption, clinoptilolite ammonia exchange columns, or reverse osmosis) and as a final polishing step following other processes.

Granular media filtration involves passage of water through a bed of granular material with resulting deposition of solids. Eventually the pressure drop across the bed becomes excessive or the ability of the bed to remove suspended solids is impaired. Cleaning is then necessary to restore operating head and effluent quality to acceptable levels. Most filters operate on a batch basis, the entire unit being removed from service for periodic cleaning.

The time in service between cleanings is termed the run length. The head loss at which filtration is interrupted for cleaning is called the terminal head loss.

9.1.2 General Design Considerations

Filter design involves selection of the following filter characteristics (1):

1. Filter configuration

2. Media sizes and depths and materials

3. Filtration rate (gpm/sq ft)

4. Terminal head loss (ft of water)

5. Method of flow control

6. Backwashing design features

The major goal in design is to achieve effluent quality objectives at low capital and operating costs. The most important characteristic in determining capital costs is the filtration rate, which fixes the filter size. Operating costs are affected primarily by filtration rate, terminal head loss, media characteristics and backwash design. The first three filter characteristics determine the cost of power for operating head and the production of the filter per run. The backwash design determines the cost per cleaning of operator attention, washwater pumping, air scouring (compressor operation) and treatment of dirty washwater. The cost of cleaning per unit volume treated (cost per cleaning divided by production per run) depends on all four factors.

Section 9.3 discusses the interrelation and effects on performance of process variables describing the characteristics of filters and of influent wastewaters. Many investigators have attempted to relate filter performance quantitatively to these variables (2) (3) (4) (5) (6). Unfortunately, such relations are of little help in predicting performance except under specific conditions already explored in pilot work. In part, this is due to the wide variations in the filtering characteristics of wastewater solids and to the dearth of reproducible objective data from well-conceived studies of wastewater filtration. In part, however, it may be inherent in the nature of the filtration process that any fully general quantitative relations describing it would be too complex for practical use. Nevertheless existing theoretical relations are useful in providing: 1) general insight into filter behavior, 2) frameworks for analysis of data from pilot investigations and 3) bases for comparing cost effects of alternatives in specific applications.

9.1.3 Basis for Design

Wherever possible, designs should be based on pilot filtration studies of the actual waste (Section 9.9). Such studies are the only way to assure:

1. Meaningful cost comparisons between different filter designs capable of *equivalent performance* (7), i.e., producing the same output quantity and quality over the same time period.

2. Most economical selection of filter rate, terminal head loss and run length for a given media application.

3. Definite effluent quality performance for a given media application.

Pilot studies are also useful for determining effects of pretreatment variations or for characterizing *filterability* in terms of performance attainable with a specific filter design. Where there is no opportunity for pilot studies, parameters for workable designs can still be determined from the discussions of wastewater and filter characteristics in the sections below. The parameters will necessarily be conservative and will tend to give more costly designs and less assurance of effluent quality than parameters based on testing. Facilities designed without pilot testing are likely to be small ones, for which the design should provide long filter runs and minimize required operating attention.

Another approach to obtaining economical facilities is to prepare a functional specification which will permit competitive bidding between suppliers of alternative filter systems. Functional requirements should include:

1. Guarantees of specified performance; both capacity and effluent quality.
2. Guarantees of proposed values of all factors which affect operating costs such as head or power requirements and backwash volume to be recycled.

Bids should be evaluated based on total present worth including operating costs, which should be calculated by a predetermined formula using factors in the guarantee.

This approach will work best when bidders are supplied test results characterizing the filterability of the waste flow. In any case, bidders should be given full information on the wastewater and treatment to be provided ahead of filtration plus the opportunity of testing effluents from such treatment where already in operation.

9.2 Process Alternatives

Filter units generally consist of a containing vessel, the granular media, structures to support or retain the media, distribution and collection devices for influent, effluent and washwater flows, supplemental cleaning devices, and necessary controls for flows, water levels or pressures.

Some of the more significant alternatives in filter layout are discussed below.

9.2.1 Alternative Flow Directions

Most filter designs employ a static bed with vertical flow either downward or upward through the bed. The downflow designs traditionally used in potable water treatment (Figures 9-1 and 9-2 (a) (d) and (e)) are most common, but recently a number of installations have been designed for upward flow (Figures 9-3 and 9-2 (b)). The European biflow design (Figure 9-2 (c)) employs both flow directions with the effluent withdrawn from the interior of the bed. Upflow washing is used regardless of the operating flow direction. Two special filter designs employ horizontal radial flow through an annular bed. Media is cycled downward through the bed, withdrawn at the bottom, externally washed, and returned to the top. (See Section 9. 10).

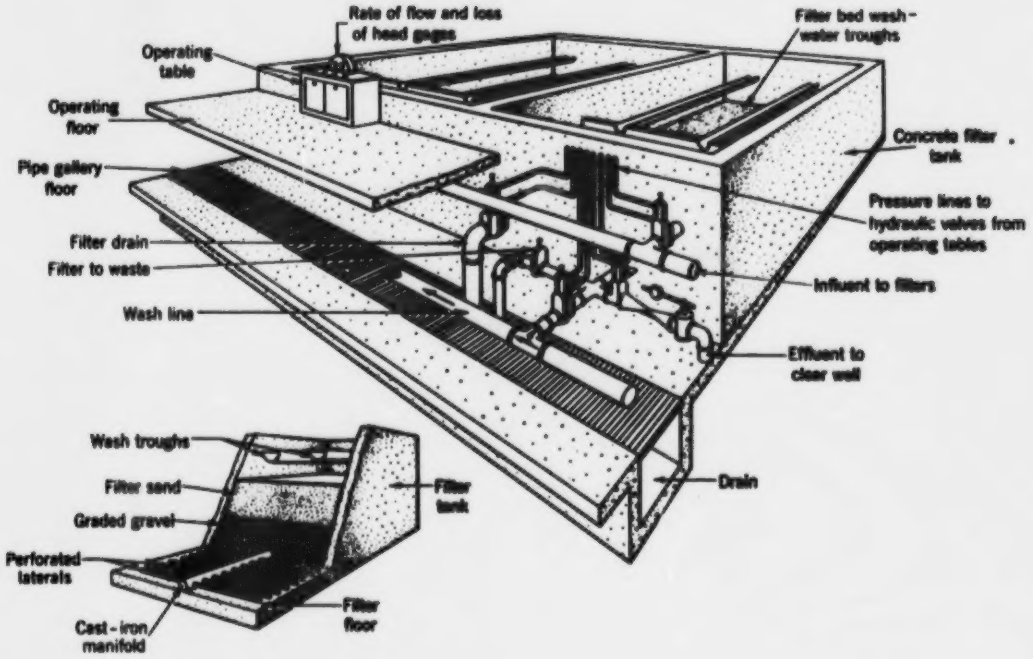
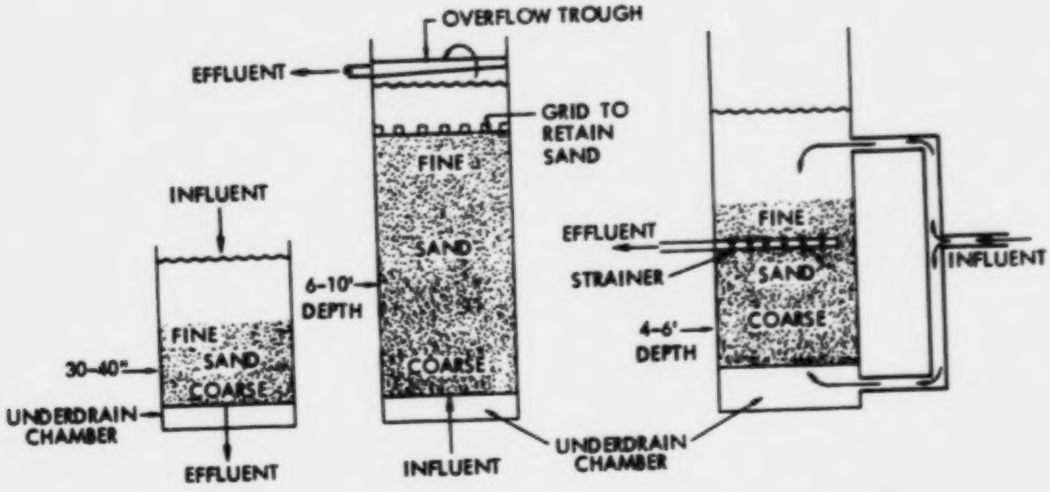
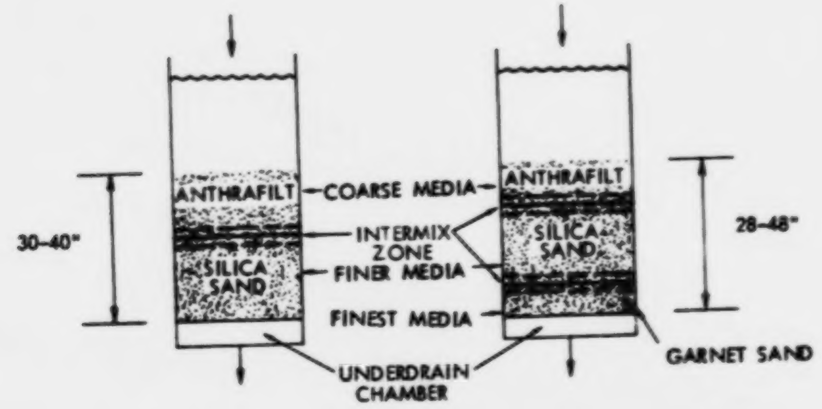

Rate of flow and loss of head gages
Filter bed wash-water troughs
Operating table
Operating floor
Concrete filter tank
Pipe gallery floor
Pressure lines to hydraulic valves from operating tables
Filter drain
Filter to waste
Influent to filters
Wash line
Effluent to clear well
Wash troughs
Filter sand
Filter tank
Drain
Graded gravel
Perforated laterals
Filter floor
Cast-iron manifold

FIGURE 9-1
TYPICAL RAPID SAND FILTER


OVERFLOW TROUGH
EFFLUENT
GRID TO RETAIN SAND
FINE
SAND
COARSE
6-10' DEPTH
INFLUENT
FINE
SAND
COARSE
30-40"
UNDERDRAIN CHAMBER
EFFLUENT
INFLUENT
UNDERDRAIN CHAMBER
EFFLUENT
STRAINER
FINE
SAND
COARSE
4-6' DEPTH
INFLUENT

(a), CONVENTIONAL FILTER (b), UPFLOW FILTER (c), BI-FLOW FILTER

SINGLE MEDIA FILTERS


30-40"
ANTHRAFILT
SILICA SAND
COARSE MEDIA
INTERMIX ZONE
FINER MEDIA
FINEST MEDIA
UNDERDRAIN CHAMBER
ANTHRAFILT
SILICA SAND
28-48"
GARNET SAND

(d), DUAL MEDIA FILTERS

(e), MIXED-MEDIA FILTERS
(TRIPLE MEDIA)

FIGURE 9-2

FILTER CONFIGURATIONS

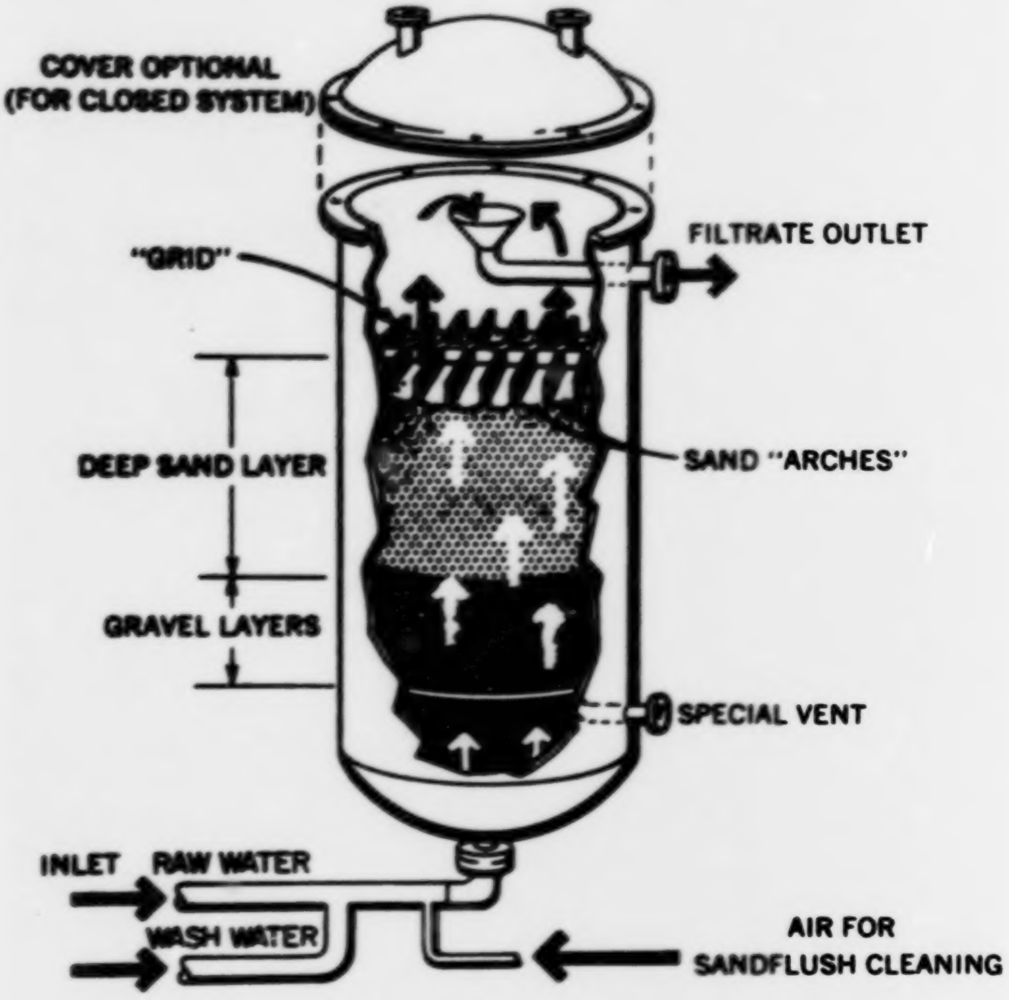

COVER OPTIONAL
(FOR CLOSED SYSTEM)
"GRID"
FILTRATE OUTLET
DEEP SAND LAYER
SAND "ARCHES"
GRAVEL LAYERS
SPECIAL VENT
INLET
RAW WATER
WASH WATER
AIR FOR
SANDFLUSH CLEANING

FIGURE 9-3

CROSS SECTION OF UPFLOW FILTER

9.2.2 Gravity vs. Pressure Filtration

Filters may be designed with closed vessels permitting influent pressures above atmospheric (Figure 9-4) or with open vessels where only the hydrostatic pressure over the bed is available to overcome filter headlosses (Figure 9-1). Pressure units are generally preferable where high terminal headlosses are expected or where the additional head will permit flow to pass through downstream units without repumping (1) (8). They are most commonly used in small-to-medium-sized treatment plants where steel-shell package units are economical (8).

9.2.3 Media Alternatives

Figure 9-2 shows schematically a number of different filter configurations using fixed bed media. The beds shown are all graded during upflow washing so that the finer material of a given specific gravity is on top. It should be noted that the conventional single media filter used in potable water treatment (Figure 9-2(a)) is generally unsatisfactory for wastewater treatment because the wastewater solids cause a high headloss buildup at the fine surface layer.

In upflow designs, flow passes first through the coarser media which for a given head loss buildup has greater capacity for retaining filtered solids. This is advantageous in lengthening filter runs and increasing output. Dual and multi-media (Figure 9-2 (d) and (e)) obtain the same effect under downflow operation by placing coarser layers of lighter material over finer denser material. An alternative downflow single-media configuration not shown attempts to get the same advantage from use of beds of uniform-sized coarse media with depths of 60 in. or more. The effects of significant media characteristics such as size gradation, specific gravity and depth, on filter performance are discussed in detail in Section 9.3.

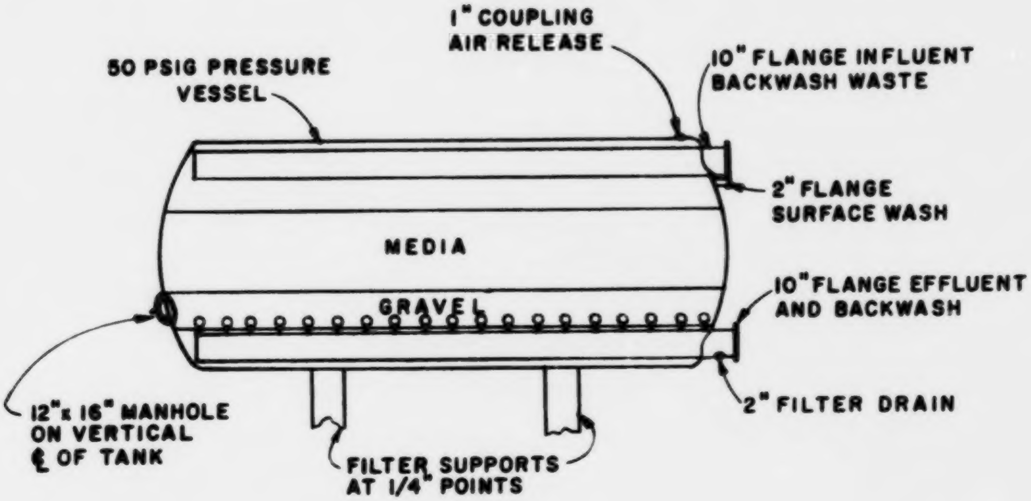
In filters using external wash, the media is not vertically graded; particle size distribution tends to be the same throughout the bed.

9.2.4 Batch vs. Continuous Operation

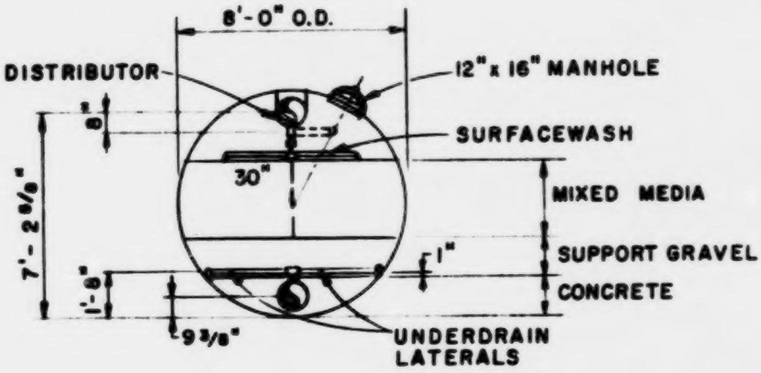
It is normal practice to design filters to operate on a batch basis with entire units taken out of service for cleaning according to schedule or as required. Several special designs, however, provide more or less continuous cleaning, either externally with media cycled through the bed, or in-place with techniques such as traveling backwash or air pulsing of the bed and air mixing of the liquid above it.

9.2.5 Normal Design

Most of this chapter will relate to fixed bed systems with intermittent upflow washing. Both upflow and downflow designs will be included as normal design. Proprietary designs using non-fixed beds or special washing systems are discussed in Section 9.10.


1" COUPLING
AIR RELEASE
50 PSIG PRESSURE
VESSEL
10" FLANGE INFLUENT
BACKWASH WASTE
2" FLANGE
SURFACE WASH
MEDIA
GRAVEL
10" FLANGE EFFLUENT
AND BACKWASH
2" FILTER DRAIN
12" x 16" MANHOLE
ON VERTICAL
℄ OF TANK
FILTER SUPPORTS
AT 1/4" POINTS

ELEVATION


8'-0" O.D.
DISTRIBUTOR
12" x 16" MANHOLE
8"
SURFACEWASH
30"
7'-2 5/8"
MIXED MEDIA
SUPPORT GRAVEL
1"
1'-6"
CONCRETE
9 3/8"
UNDERDRAIN
LATERALS

SECTION

FIGURE 9-4

TYPICAL PRESSURE FILTER
(Courtesy of Neptune Microfloc, Inc.)

9.3 Process Variables

9.3.1 Performance Relations

The measures of filter performance are output quality and quantity. The variables which determine or limit performance fall into two major groups: influent characteristics and the physical characteristics of the filter. The latter include media characteristics, filtration rate, available and applied operating head and the design and operating parameters of the filter cleaning system.

In determining the fundamental limits on quality performance,the characteristics of prime importance are those of the influent solids: concentration, strength, size, and the physical-chemical properties governing adhesion of particles to each other or to the media surfaces. Commonly a number of filters with different physical characteristics can come close to the limiting quality performance for a given influent. In contrast, quality performance of given filters can vary widely for different solids characteristics.

In determining output quantity from filters, the influent solids characteristics especially floc strength and solids concentration are again very inportant, but the physical characteristics of the filter become significant too.

At run lengths of 24 hr or more, output depends almost totally on filter rate. As run lengths become shorter, however, the effects of downtime and washwater recycle during cleaning become increasingly important (See Section 9.4). The washwater recycle volume depends on the backwash flow rates and the wash cycle duration needed for adequate cleaning (see Section 9.7). Factors governing backwash system design include:

1. Size distribution, depth and specific gravity of media
2. Nature of solids removed, principally their adhesion to the media and their tendency to compact in a dense layer at the media surface
3. Type of supplementary cleaning provided.

Run length may be limited either by available head or by deterioration of effluent quality as the filter bed becomes filled with solids (breakthrough). Which factor governs depends on the interaction of several variables including:

1. Influent solids characteristics (all those which affect quality performance)
2. Flow rate
3. Temperature and viscosity of the wastewater
4. Media characteristics
5. The amount of head available.

Headloss in a clean bed varies directly with filter rate and inversely with grain size. In determining head loss buildup, the most significant media characteristic is the grain or pore size at the influent surface of the bed (or in some cases within finer denser layers of multi-media filters). In downflow filtration through a graded bed, influent solids particles larger than about 7 percent of the minimum grain size (9) will be removed by straining provided their strength is sufficient to withstand the shear at the surface. Shear varies with filter rate and liquid viscosity.

In surface straining, head loss increases exponentially with time or solids accumulation (See Chapter 8). Where significant solids loads are removed predominantly by surface straining, head loss buildup will be rapid, filter runs short and backwash frequency high. In addition the solids removed at the surface tend to be compressed into a dense mat which is difficult to remove in backwashing.

Removal of solids within the bed rather than just at the surface is termed depth filtration. Both surface and depth filtration are usually involved to some degree in any given application.

In depth filtration head loss tends to build up linearly with time or with solids accumulation. Compression of the solids removed is limited by the granular structure of the bed. For downflow filtration within a single media, the farther solids penetrate into the bed, the slower will be the rate of head loss buildup, but the sooner solids will breakthrough into the effluent.

The factors which determine breakthrough for a single media are the media size and depth, the flow rate and the resistance of deposited materials to shear within the bed. Hudson (10) suggested characterizing the resistance of solids to breakthrough by an index, K, calculated from the physical characteristics of the filter and the head loss at which breakthrough occurs. The expression for the index is:

$$K = Vd^3H/L$$

Where:

V = filtration rate—gpm/sq ft
H = head loss at breakthrough—ft
d = effective size of media—mm
L = bed depth—ft

9.3.2 Influent Characteristics

The influent characteristics of prime importance in determining filter performance are those of the solids to be removed. The only significant characteristic of the wastewater liquid—as opposed to the solids—is viscosity which varies with temperature. Its effects on development of filter head loss are generally small, however, in comparison to the effects of solids accumulations or filter rates.

The characteristics of wastewater solids which govern or limit filter performance are determined by the treatment processes ahead of filtration (see Section 9.1). In direct filtration of secondary biological effluent the residual solids applied to the filter are predominantly biological floc grown in the treatment process. In filtration of effluent following tertiary coagulation for phosphate removal the residual solids are predominantly chemical flocs. In filtration of chemically precipitated raw wastewater or primary effluent, the solids consist of inorganic chemical floc with varying quantities of precipitated organics.

Loading, media and performance data for filter applications of the above three types are shown in Tables 9-1, 9-2 and 9-3. Most data are for full scale installations but a few large pilot facilities are included. The data are those ordinarily recorded in tests of filter installations. These data show only that the systems filtering physical-chemical floc tend to use somewhat lower filter rates and somewhat finer media (with dual or multi-media configurations almost standard) and that effluent results for any given type of influent source may vary considerably.

Compiling data from a number of filter installations treating biological effluents, Kreissl (33) found removals ranging from 50 to 90 percent with a mean of about 70 percent, provided influent solids were below 35 mg l. Included were data for a variety of loading rates media configurations and types of prior biological treatment. Subsequent compilation of similar data for effluents from chemical treatment systems showed mean solids removals of only 60 percent (34), indicating that on the average chemical floc tends to be more difficult to filter.

The only influent solids characteristic included in routine filter testing is the concentration, perhaps because it is the only one that is easily measured. A few special studies have attempted to take into account other characteristics such as floc strength, particle size distribution (concentration vs. size) and properties governing adhesion of particles to each other or to the media. Some other studies have tried to distinguish differences in filter performance according to parameters of the treatment prior to filtration. Outlined below are a few significant additional insights into wastewater filtration provided by these special studies.

9.3.2.1 Floc Strength

Biological flocs tend to be significantly stronger or more resistant to shear than chemical flocs, at least those from alum or iron coagulants (2). Consequently, in filtering biological flocs, surface straining is generally significant and runs are almost always terminated by excessive head loss. Breakthrough is rarely observed. In one study head losses as high as 30 ft were applied without deterioration of effluent quality (35). This contrasts with alum and iron (hydroxide) flocs which have been shown to penetrate readily into filters and to breakthrough at relatively low heads ranging from 3 to 6 ft (2) (10) (36). In an isolated instance where breakthrough of biological floc was observed, the index K was found to be 13.7, far above the range of 0.3 to 3.6 cited for alum or iron floc in water treatment (37). In contrast to flocs from other common coagulants, calcium carbonate precipitates are strongly removed at the filter surface where they may form a dense compressed layer hard

TABLE 9-1

RESULTS OF STUDIES OF FILTRATION OF EFFLUENT FROM SECONDARY BIOLOGICAL TREATMENT

LOCATION	TYPE OF FILTER	INFLUENT SOURCE	BED CHARACTERISTICS			HYDRAULIC LOADING	SUSPENDED SOLIDS			RUN LENGTH	REFERENCE
			MEDIA	SIZE	DEPTH		IN	OUT	REMOVAL		
			type	mm	in	gpm/ft^2	mg/l	mg/l	percent	hr	
Luton, G.B.	Gravity Downflow	Activated Sludge	Sand		36	1.6-4.0	25-50	3-6	72-91	12	(11)
	Gravity Downflow	Trickling Filter	Sand		36	3.4	28-35	9-10	67-74	-	(12)
						5.0	13	8	40	-	
Hanover Park, Illinois	Gravity Downflow	Activated Sludge	Coal	1.2-1.3	30	2.0	14	4	57	106	(13)
			Sand	.8- .9	12	4.0	16	4	67	27	
			Garnet	.4- .8	6						
	Pressure Upflow	Activated Sludge	Sand	1-2	60	2.0	14	7	50	150	(13)
						4.0	15	6	67	17	
						5.0	13	6	54	7	(14)
	Pressure Downflow	Activated Sludge	Coal	1.4-1.8	24	2	16	7	56	90	
			Sand	0.8-1.0	12	4	15	5	67	15	
						6	16	6	62	22	
						8	13	6	54	31	
						10	18	8	55	12	
Walled Lake-Novi, Michigan	Gravity Downflow	Activated Sludge	Mixed media	0.25-2.0	30	3-4	7	3	57		(15)
Louisville, Ky. (Hite Creek)	Pressure Horizontal	Activated Sludge	Coal	1.0-1.2	16.5	3.4	27	3	89	-	(16)
			Sand	0.45-0.55	9						
			Garnet	0.2-0.3	4.5						
Coldwater, Michigan	Pressure Downflow	Trickling Filter	Coal	0.8	20	4.9	21	8	62	2.5-8	(15)
			Garnet	0.4-0.6	20						
			Garnet	1.2	9						

TABLE 9-1 (CONTINUED)

RESULTS OF STUDIES OF

FILTRATION OF EFFLUENT FROM SECONDARY BIOLOGICAL TREATMENT

LOCATION	TYPE OF FILTER	INFLUENT SOURCE	BED CHARACTERISTICS			HYDRAULIC LOADING	SUSPENDED SOLIDS			RUN LENGTH	REFERENCE
			MEDIA	SIZE	DEPTH		IN	OUT	REMOVAL		
			type	mm	in	gpm/ft^2	mg/l	mg/l	percent	hr	
Bedford Township, Michigan	Pressure Downflow	Activated Sludge	Multi-media	-	-	-	15	3	80	15	(15)
Ventura, California	Gravity Deep Bed Downflow	Trickling Filter	Sand	1-2	-	6	18	7	61	6-18	(17)
West Hertfordshire, G.B.	Immedium Upflow Pressure	Activated Sludge	Gravel Sand	- 1-2	26 60	2.2 4.0 5.0 6.0	44 37 55 37	2 4 7 10	95 90 87 73	- - - -	(18)
Ann Arbor, Michigan	Downflow	Activated Sludge	Multi-media	-	-	6	42	5	88	-	(19)
State College, Pa.	Pressure Downflow	Activated Sludge	Sand	-	84	3-12	6	1	85	6	(20)
Springfield, Ohio	Gravity Downflow	Contact Stabilization	Sand	0.45	10	5.3	14	5	64	-	(21)

TABLE 9-1 (CONTINUED)

RESULTS OF STUDIES OF FILTRATION OF EFFLUENTS FROM SECONDARY BIOLOGICAL TREATMENT

LOCATION	TYPE OF FILTER	INFLUENT SOURCE	BED CHARACTERISTICS			HYDRAULIC LOADING	SUSPENDED SOLIDS			RUN LENGTH	REFERENCE
			MEDIA	SIZE	DEPTH		IN	OUT	REMOVAL		
			type	mm	in	gpm/ft^2	mg/1	mg/1	percent	hr	
Letchworth, England	Pressure Upflow	Activated Sludge	Sand	1-2	60	5.3	17	7	60	-	(22)
Upper Stour Main Drainage, Freehold Works, England	Gravity Downflow	Activated Sludge	Sand	0.5-2.5	-	1.2-2.4	12	5	58	-	(22)
Harpenden, U.D.C.	Gravity Downflow	Trickling Filter	Sand	1.1	-	1-3	20	5	75	-	(22)
Rodbourne Works, Swindon, England	Gravity Downflow	Trickling Filter	Sand	1.5-3	-	1.6-3.2	21	5	75	-	(22)
Derby, England	Simater Radial Flow	Trickling Filter	Sand	1-2	-	4-6	22	9	60	-	(22)
Thameside, England	Immedium Pressure Upflow	Activated Sludge	Sand	1-3	63	3.3	9	2	74	-	(23)
						3.3	46	8	84		
						5.0	8	6	20		
						5.0	37	10	74		
Thameside, England	Permutit Upflow	Activated Sludge	Sand	0.60-1.20	57	3.3	9	1	86		(23)
						3.3	32	7	78		
						5.0	11	4	60		
						5.0	28	5	83		
Thameside, England	Simater Radial Flow	Activated Sludge	Sand	0.5-1		3.3	11	3	74		(23)
						3.3	51	7	86		
						5.0	11	4	62		
						5.0	24	10	58		
Ashton-Under-Lyne, England	Immedium Upflow	Trickling Filter	Sand	1-2	60	4.5-5.0	30	8	80		(24)

TABLE 9-2

RESULTS OF STUDIES OF FILTRATION OF CHEMICALLY TREATED SECONDARY EFFLUENT

LOCATION	TYPE OF FILTER	INFLUENT SOURCE	BED CHARACTERISTICS			HYDRAULIC LOADING	SUSPENDED SOLIDS			RUN LENGTH	REFERENCE
			MEDIA	SIZE	DEPTH		IN	OUT	REMOVAL		
			type	mm	in	gpm/ft^2	mg/l	mg/l	percent	hr	
Piscataway, Md.	Pressure Downflow	A.S.+2-Stage Lime Clarification	Coal Sand	1.0 0.5	12 6	3	12	8	33	50	(26)
Ely, Minnesota	Gravity Downflow	High Rate +2-Stage Lime Clarification	Coal Sand		24 12	2.3	8	<2	>75	24	(27)
Jefferson Parish, La.	Upflow	T.F.+In-Line Alum Injection	Sand		-	3	40	21	48	2.5-6.5	(28)
Nassau County, N.Y.	Gravity Downflow	A.S.+Alum Clarification	Coal Sand	0.9 min. 0.35 min.	36 12	2.5-3.5	2-10	0-2	80-90	16-48	(29)
Lake Tahoe, California	Two-Stage Pressure	A.S.+Lime Clarification+Ammonia Stripping and Recarbonation	Coal Sand Garnet		18 12 6	2.8-4.0	9-15	0-1	93	4-60	(30)

TABLE 9-3

RESULTS OF STUDIES OF FILTRATION FOLLOWING CHEMICAL TREATMENT OF PRIMARY OR RAW WASTEWATER

LOCATION	TYPE OF FILTER	INFLUENT SOURCE	BED CHARACTERISTICS			HYDRAULIC LOADING	SUSPENDED SOLIDS			RUN LENGTH	REFERENCE
			MEDIA	SIZE	DEPTH		IN	OUT	REMOVAL		
			type	mm	in	gpm/ft^2	mg/l	mg/l	percent	hr	
Washington, D.C.	Gravity Downflow	Two-Stage Lime Clarification	Coal Sand	0.9 0.45	18 6	1.7-6.3	14	6	70	12-50	(32)
Lebanon, Ohio	Gravity Downflow	Single Stage Lime Clarification	Coal Sand	.75 .46	18 6	2.0	30	10	67	-	(31)
Washington, D.C.	Gravity Downflow	Two-Stage Lime Clarification	Coal Sand	1.2-1.4 0.6-0.7	24 12	2.4-4.4	139	33	74	>24	(25)
Washington, D.C.	Gravity Downflow	Single Stage Lime Clarification	Coal Sand	1.2-1.4 0.6-0.7	24 12	2.3-4.3	123	23	81	>24	(25)

to remove during washing (36). Comparative data are lacking on the strength of flocs from precipitation of phosphates in wastewater using alum, iron or lime. It is reasonable to assume, however, that they are similar to aluminum or ferric hydroxide floc.

Polymer filter aids may be added to the filter influent to strengthen weak chemical flocs thereby permitting operation at higher rates without breakthrough. Doses of 0.1 mg/l or less are often adequate (8). Polymers added as coagulant aids in upstream settling or flocculating units may similarly strengthen the residual floc applied to the filters. Ample head loss must be available to meet losses due to the tougher floc, and doses must be kept as low as possible to avoid excessive head loss.

9.3.2.2 Particle Size

Floc particle sizes in settled biological effluent tend to be bimodally distributed. Mean sizes for the two modes in one study were 3 to 5 microns and 80 to 90 microns (2). About half of the weight was in each mode. Theoretical work (38) suggests that particles in the lower size range are much less effectively removed by filtration than those in the higher range. Hence for the best quality performance from filtration, the proportion of smaller size particles must be reduced to a minimum by proper flocculation.

9.3.2.3 Filterability

The filterability of residual solids from secondary settling varies with solids retention time and with liquid contact time in the biological process. For biological systems with higher solids retention times and longer liquid contact times, filtered effluents tend to have lower suspended solids. Culp and Culp (8) indicated the expected performance of multi-media filters for plain filtration in secondary effluents as shown in Table 9-4.

TABLE 9-4

EXPECTED EFFLUENT SUSPENDED SOLIDS FROM MULTI-MEDIA FILTRATION OF SECONDARY EFFLUENT

Effluent Type	Effluent SS mg/l
High Rate Trickling Filter	10-20
2-Stage Trickling Filter	6-15
Contact Stabilization	6-15
Conventional Activated Sludge	3-10
Extended Aeration	1-5

It is significant that the solids in extended aeration effluents filter particularly well, in as much as they often settle poorly, leaving high concentrations in the secondary effluent. This behavior may be understood from the flocculation studies of Parker, et al, (39) who found that sludges with high solids retention times lose their tendency to agglomerate into larger easily settleable particles, but increase in strength so that fewer are broken up into particles of a size not readily filtered.

9.3.2.4 Headloss Buildup vs. Solids Capture

While effluent quality reflects the solids which pass through the filters, headloss development reflects the amount and location of solids which deposit in the bed. Both solids loading (solids concentration times flow rate) and filter efficiency are important in determining the buildup of headloss with increasing solids capture. Various studies relating headloss buildup to solids capture show widely different results. This would be expected in view of the wide range of solids characteristics, media characteristics and filter rates, and the very different headloss patterns that result from surface and depth filtration. Baumann and Cleasby (1) cite specific solids capture values (average over the filter run) ranging from 0.035 to 0.35 lb/sq ft/ft of headloss. The variation was mainly in activated sludge effluent. The trickling filter data, from a single plant in Ames, Iowa, showed values close to 0.07 lb/sq ft/ft of headloss for a wide range of media sizes and filter rates. British data for trickling filter effluent, however, showed specific capture values averaging 0.35 lb/sq ft/ft of headloss over a filter run with initial values as high as 0.6 lb/sq ft/ft of headloss (40). For a fine (0.5 mm) media with low solids loadings, Tchobanoglous and Eliassien (2) reported values an order of magnitude lower than the smallest cited by Baumann and Cleasby. It is reasonable to expect the highest values of specific capture where the filter and influent solids characteristics permit depth filtration and extremely low values where they promote a high degree of surface straining.

9.3.2.5 Properties of Solids Affecting Adhesion

Available measures of the properties which affect adhesion of solids particles to other solids or to media grains are limited to Zeta potential or the related electrophoretic mobility. Very few studies have included such measures or attempted to relate performance to them.

Tchobanoglous (6) reported that reduction of natural negative electrophoretic mobility of wastewater solids using cationic polymers improved filter quality performance. Where sufficient polymer was added to reverse the negative charge of the particles performance, though excellent at first, deteriorated rapidly after the first hour. With charge reversal, initial performance apparently was aided by electrostatic attraction between the negatively charged filter media grains and the positively charged wastewater solids. After an hour, however, the grains were coated with positively charged solids, and the resulting electrostatic repulsion interfered with filtration of further solids applied.

9.3.3 Physical Characteristics of the Filter

Most wastewater filter designs employ media configurations and loadings which minimize surface straining and promote depth filtration (Section 9.4). A few special designs with fine media (Section 9.6) are intended to remove solids primarily by surface filtration or straining. These designs include provisions for overcoming the adverse effects of rapid headloss buildup. Where surface filtration predominates, the media characteristics have little effect on quality performance or head loss. In addition, removal of solids is quite independent of filter rate or influent solids concentration (1). Hence the effects of physical characteristics of filters are discussed below only in relation to depth filtration not surface straining.

9.3.3.1 Media Characteristics

The most important media characteristic in determining performance is size. Studies using uni-size media have clearly demonstrated that finer media have greater removal efficiency (2) (6) (7) (35) (41). Various investigators have related percent removal to powers of diameter ranging from -1 to -3 (3). In finer media headloss per unit of removal (lb/cu in. of media) is also higher (2).

In a media graded from fine to coarse in the direction of flow, the highest solids concentration is applied to the layers with the greatest removal efficiency. As a result, removal is concentrated in a small depth with accompanying high headlosses.

In media graded from coarse to fine in the direction of flow, substantial penetration occurs but most of the solids are removed in the coarser media where less head loss buildup results. The finer layers, protected from heavy solids loadings, are available for polishing and to prevent breakthrough as the coarser layers become filled with solids.

Media depth is most significant in coarse uniform beds. Because of the uniformity, the efficiency of removal (as a percent of the solids applied to each depth) is nearly constant for all layers of the filter. Penetration is substantial and extra depth is relied upon for polishing and to retard breakthrough.

Size and specific gravity of media together are significant in determining expansion during backwash and the degree of intermixing in multi-media beds.

9.3.3.2 Filter Rates

The effect of filter rates on quality performance can vary widely depending on application. In filtering biological floc at reasonably low influent solids concentration, the effect on effluent quality of rates up to 10 gpm/sq ft is not very significant (24) (37) (42). In a study of ultra high rate filtration (43), operation at up to 32 gpm/sq ft still provided 50 percent removals compared to 75 percent at 8 gpm/sq ft. With weaker chemical flocs or with high influent concentrations of biological solids (usually indicating poorly functioning biological treatment) filter effluent quality tends to degrade at filter rates above about 5 gpm/sq ft (33). Sudden changes in filter rates may affect effluent quality more adversely than sustained higher rates.

Higher filter rates tend to increase solids penetration. In cases where this significantly reduces surface removal, head loss buildup per unit volume filtered may actually be less at higher rates. This was illustrated in studies at Iowa State University involving settled trickling-filter and lime-softening effluents (35) (36). For the trickling filter effluent, production per run (at a given terminal head loss) increased slightly with filter rate over the range tested up to 6 gpm/sq ft. For the lime effluent, production per run increased with filter rate up to 5 gpm/sq ft and then decreased (36). Existence of an optimum rate, as in the latter study, has been suggested as typical of combined surface and depth filtration (1). It has also been suggested that the advantages of using a coarse top media layer may be lost if the filter rate is not high enough to force solids into the bed and limit surface straining (7).

9.3.3.3 Cleaning System Variables

In addition to upflow washing, some form of auxiliary scouring of the media appears essential to adequately clean wastewater filters. If cleaning is inadequate, two serious problems will develop: filter bed cracking and mud ball formation. Cracks open in filter beds because of compression of excessively thick coatings on the filter grains. The resulting localized heavy penetration of solids may both lower effluent quality and contribute to mud ball formation.

Mud balls are compressed masses of filtered solids large and dense enough to remain in the bed during backwashing. If conditions favoring their formation persist, mud balls tend to increase in size and to sink deeper in the bed. Their presence increases head loss and may lead to loss of effluent quality.

Both air scrubbing and surface or internal water jets have been used for auxiliary scouring of the media. Air injected below the media produces shear as the bubbles rise through the bed. Water jets, positioned at the top of the expanded bed, produce high shear around the surface media, which is the most heavily loaded with solids. In multi-media beds, jets may be similarly provided at the expanded height of the media interface.

The main upflow wash and the auxiliary scouring systems should be controlled independently to permit use together or separately. Washing procedures are discussed in Section 9.7. The key parameters for design of the cleaning system are the upflow wash rate capacity and the air scour rate or surface wash rate capacity. Typically, upflow wash rates are about 20 gpm/sq ft. The maximum capacity is selected to provide the desired degree of fluidization and expansion of the media under critical high temperatures (See Section 9.7). Capacities for auxiliary scouring are generally established empirically. Air scour rates typically range from 3 to 5 scfm/sq ft, and surface wash rates from 1 to 3 gpm/sq ft.

9.4 Selection of Filtration Rate and Terminal Headloss

Given adequate information on performance, the filter rate and terminal headloss for a particular media design should be selected by making economic tradeoffs between filter size, operating head requirements and run length, all within the limits dictated by effluent quality requirements. This section outlines procedures for such tradeoffs and provides an alternative basis for selection where specific performance information is lacking.

9.4.1 Information for Economic Tradeoffs

Adequate information for making economic tradeoffs can be obtained only from pilot studies of the specific media application. (See Section 9.9). Pilot studies should indicate the buildup of headloss with time for various filter rates and for average and peak influent solids concentrations. Results may be indicated in a form similar to Figure 9-5. With this information it is possible to estimate the filter run length, the net production and the capital and operating costs of the filter for the given influent solids concentrations and for different combinations of filter rate and terminal headloss (See Sections 9.1 and 9.3).

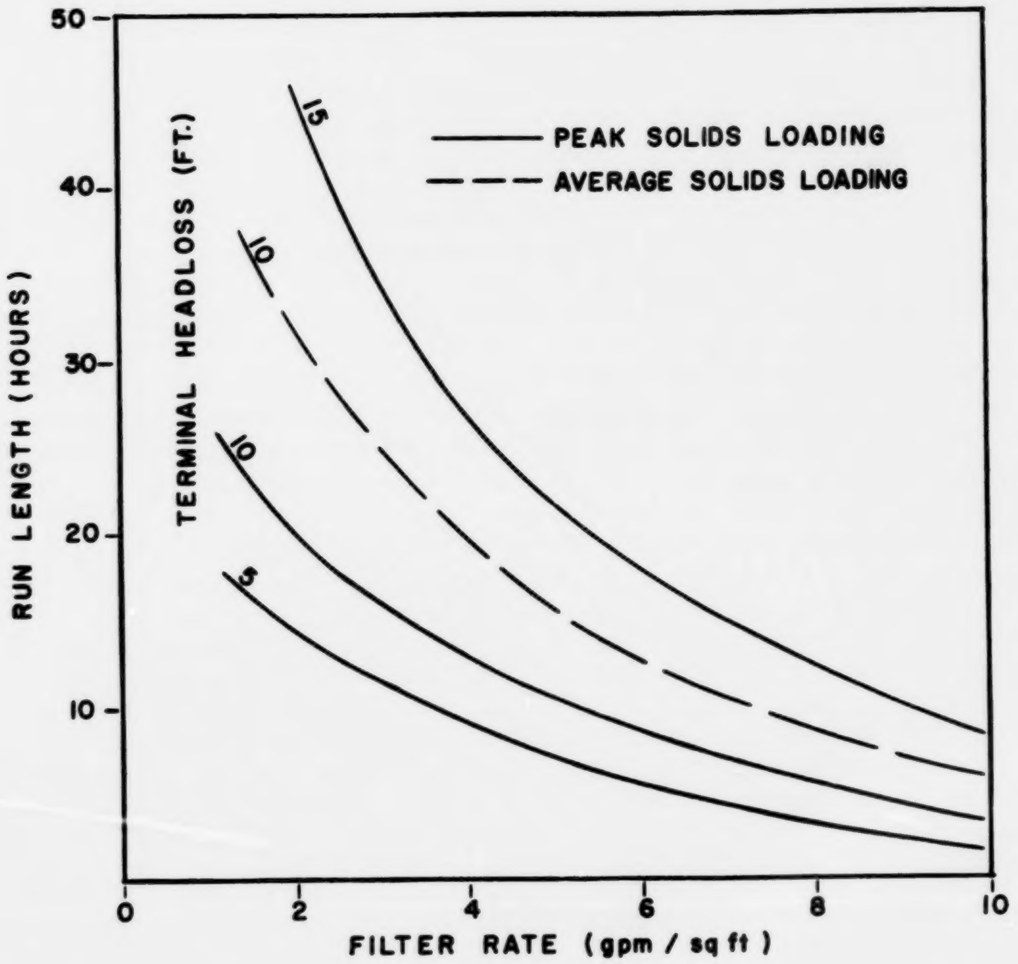


FIGURE 9-5
RUN LENGTH VS. FILTER RATE FOR
VARIOUS TERMINAL HEAD LOSSES

In determining net production, allowances must be made for downtime during cleaning and for recycle of washwater through the treatment plant. The downtime effects are calculated from the cleaning frequency, cleaning cycle duration and the number of individual filters. Washwater recycle effects are calculated from the cleaning frequency and the backwash rate and duration. Washwater recycle has no effect on net production if filter influent is used for washing. Net production may be expressed as volume (filter rate x run length) or as an average rate (gpm sq ft) over one filter cycle (run length plus cleaning time). The net production rate is almost the same as the filter rate for runs of 24-hr or more. For run lengths below 10 to 12-hr the differences become significant (1); below 6 to 8-hr the effect on production may be critical.

Short term peak loadings due to down time or recycle during backwash need not be considered directly in economic tradeoffs. After the design filtration rate and terminal headloss are determined, however, the design should be checked to assure that it can accommodate these peaks within the available headloss and effluent quality limits. If not, peak effects should be reduced or eliminated by increasing the number of filter units or by providing equalizing storage for the backwash and wastewater flows.

The design should also be checked for its ability to handle the sustained peak loads imposed when a unit is taken out of service for repairs. If the resultant shorter run lengths do not provide enough capacity, the design may be revised as follows: peak hydraulic loadings should first be reduced by increasing the number of filters keeping the total area the same; if this reduction is not sufficient, the area should be increased beyond that determined in the original design.

Before cost tradeoffs are made, the following must be defined:

1. Maxium flows and solids loadings for various durations up to 24-hr. A tentative decision is required on the use of equalization to limit maximum wastewater flows.

2. Run length limits. The lower limit should be 6 to 8 hr to maintain reasonable net production. The upper limit should be 36 to 48 hr to avoid anaerobic decomposition of solids in the filter (1).

3. Head loss limits. For gravity filters allowable head losses generally are below 10 ft. Use of heads much above this commits the design to pressure filters. Use of pressure filters would be favored where pressurized discharge to following facilities is needed (1). Gravity filters would be favored where the extra head for pressure filters would require intermediate pumping but head for the gravity units is available without such pumping.

4. Backwash design and expected cost per cycle. Manpower costs should reflect whether the operation is to be automated. Backwashing costs should include costs of treating recycled backwash in units ahead of the filters, and the recycled flow should be deducted in determining the net production of the filters.

5. Space limitations. These may force use of higher filter rates.

6. Number of filter units. This should be tentatively selected to facilitate cost estimates, but may be varied with little effect on the tradeoff calculations provided labor is not a major factor in the operating cost per backwash. For reliability and economy, a minimum of four to six units should generally be provided, with at least two in even the smallest installations. Above these minimums, the number of filter units, depends on the actual size of individual units. The practical maximum size of gravity filters is about 800 sq ft.

In addition to limits indicated above, pilot testing may reveal: 1) upper limits on headloss or rate required to avoid solids breakthrough and effluent quality deterioration, 2) an optimum filter rate for minimizing headloss buildup. No filter rates lower than the optimum should be considered in the tradeoffs.

9.4.2 Tradeoff Procedures

The following procedures are suggested for determining the most cost effective filter sizing, design terminal head loss and run length. Figure 9-6, relating net production to filter rate and run length, has been prepared to facilitate the analysis. The figure should be modified before application if backwashing conditions differ significantly from those assumed in its development.

1. From filter test data for average and maximum design influent solids concentration, prepare a headloss development plot (see example plot Figure 9-5).

2. Assume initial trial values for terminal headloss and filter run length. (See Item 12).

3. For the assumed terminal headloss and run length determine the filter rate from the headloss development plot for maximum solids concentration.

4. For this filter rate and the assumed run length determine the net production rate from Figure 9-5.

5. Determine filter sizing based on this net production rate and the maximum design flow for a duration equal to the filter cycle time (run length plus downtime for cleaning).

6. Estimate capital costs for filters based on above sizing and the design terminal headloss.

7. Determine average net production by dividing average flow by filter area.

8. Construct a plot of net production vs. filter rate based on run lenths to reach the trial value of terminal headloss at various filter rates with average solids concentra-

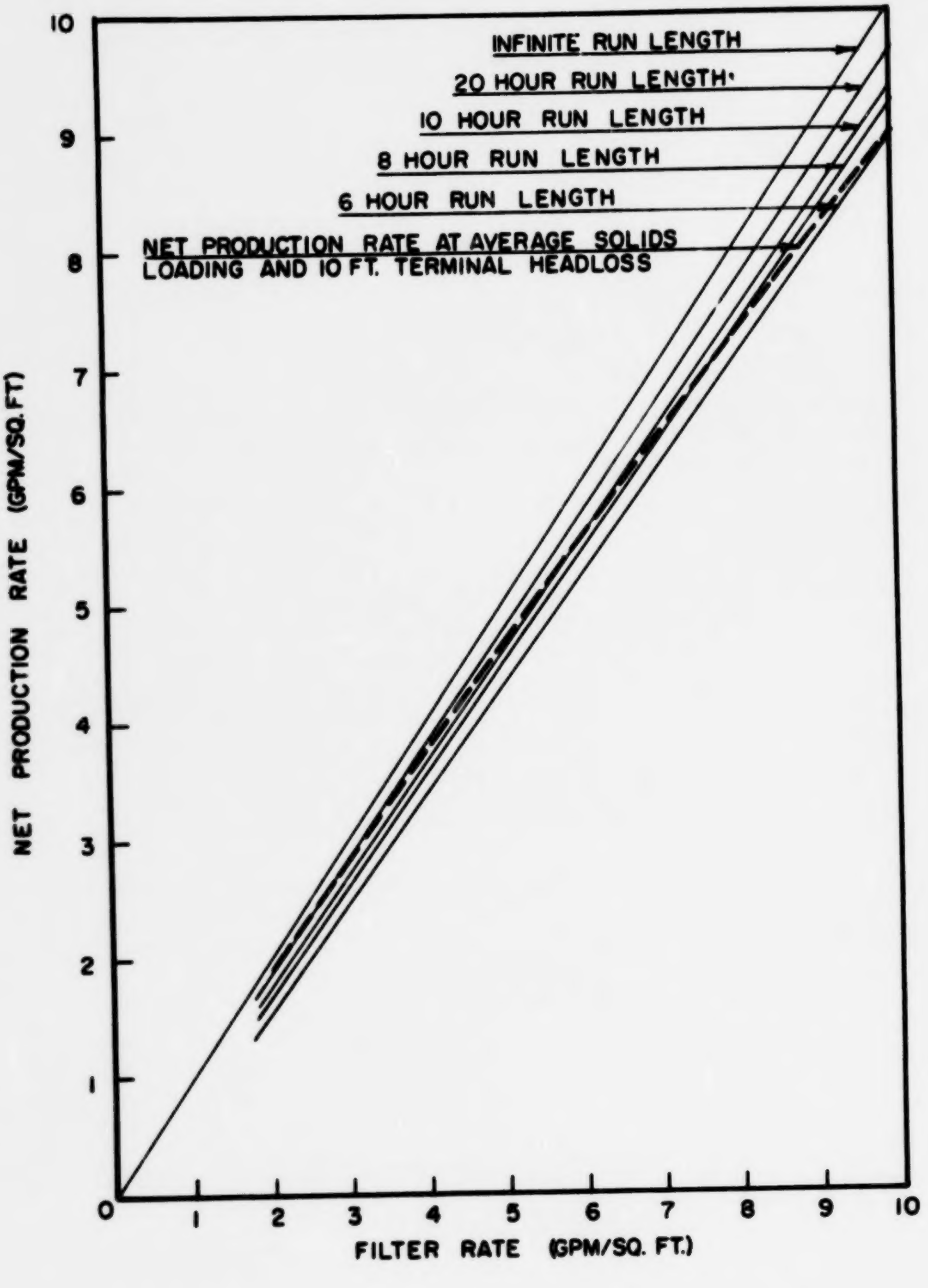


FIGURE 9-6

NET PRODUCTION RATE VS. FILTER RATE
FOR VARIOUS RUN LENGTHS

tions. (See Example Figure 9-6).

9. From the plot in 8. determine filter rate and run length to provide average net production.

10. Calculate operating costs based on the assumed terminal headloss and the run length for average flow and solids loading.

11. Convert operating costs to present worth and add to capital cost to determine total present worth.

12. Repeat above analysis assuming different values for terminal head loss and filter runs. The objective is to find assumptions which minimize present worth, within technological constraints.

    It is suggested that a conservative initial value of 8 ft be assumed for terminal headloss with a run length of at least 8 hr at maximum solids concentrations. Subsequent trials would explore use of higher headloss values to permit either longer runs or higher filter rates whichever appears more advantageous. Judgement must be applied to minimize amount of calculation required.

9.4.3 Selection Without Pilot Testing

Where it is impossible to test proposed filter media on the actual influent, guidance may be obtained from results with the same media treating similar influents. In the absence of specifically applicable test results, filter rates and headloss allowances should be very conservatively selected, based on ample estimates of influent solids concentrations. To assure adequate capacity it is suggested that, as a minimum, sufficient filter area be provided to handle the 24-hr design flow at 4 gpm/sq ft or the 4-hr maximum design flow at 6 gpm/sq ft, whichever is more stringent. For predominantly chemical floc, the surface media should be no finer than 1 mm and allowance should be made for a terminal headloss of 10 ft. For filtration of biological solids in secondary effluent the following procedures are suggested in selecting terminal headloss and final filter sizing:

1. For the minimum filter area as determined above, estimate headloss buildup based on expected solids removals and the following values of specific capture:

Minimum Media Size at Influent Surface mm	Specific Capture lb of solids removed/sq ft/ft of headloss increase
1.8	0.07
1.3	0.035

2. Avoid use of any finer surface media. Surface media coarser than 1.8 mm may permit higher specific captures but problems of adequate cleaning must be considered (See Section 9.7).

3. For the minimum filter area calculate the required head for 24-hr run length at average solids loading and for 8-hr run length at maximum (8-hr) solids loadings.

4. Provide for terminal headloss on the more critical basis above or use more than minimum filter size and recalculate solids loadings and headloss requirements.

Designs based on the criteria above should be as flexible as possible to permit use of higher rates or lower heads if operating experience shows this is possible. Flexibility to increase rates is most valuable where capacity is to be increased in future stages. Flexibility in pumping and control systems will permit head to be reduced to what proves necessary in actual operation.

9.5 Filtration Media

9.5.1 Materials

Media commonly used in water and wastewater filtration include silica sand (sp gr 2.65), anthracite coal (sp gr 1.4 to 1.6) and in special multi-media designs garnet (sp gr 4.2) or ilmenite (sp gr 4.5).

As they occur in nature these materials are not of uniform size but instead typically have a grain size distribution such as that shown in Figure 9-7. Fair and Geyer (44) discuss size measures for irregular particles, equivalent diameters, shape effects, etc. Natural grain size distributions frequently are close to geometrically normal, i.e.plot as a straight line on log probability paper. As shown in Figure 9-7, grain size distributions are often characterized by two points, the 10 percent and 60 percent size (d_{10} and d_{60}). These are sizes such that the weight of all smaller particles constitutes respectively 10 or 60 percent of the whole. Media is frequently specified in terms of effective size (d_{10}) and the uniformity coefficient (d_{60}/d_{10}).

It is possible to change the characteristics of a given media material by removing certain size fractions. Coarser fractions may be seived out while finer fractions may be removed by "scalping" (removing surface layers) after hydraulically grading material during upflow washing. Fair and Geyer (44) present a method of calculating the size fractions which must be removed to convert from one size distribution to another.

The most important modification for most media is to remove any very fine particles—say those less than 80 percent of the effective size. Such fine material never constitutes more than a small fraction of the media volume but, if not removed, may cause headlosses far greater than would be expected for the given effective size.

With sufficient effort in size separation, it is possible to produce almost uniform media.

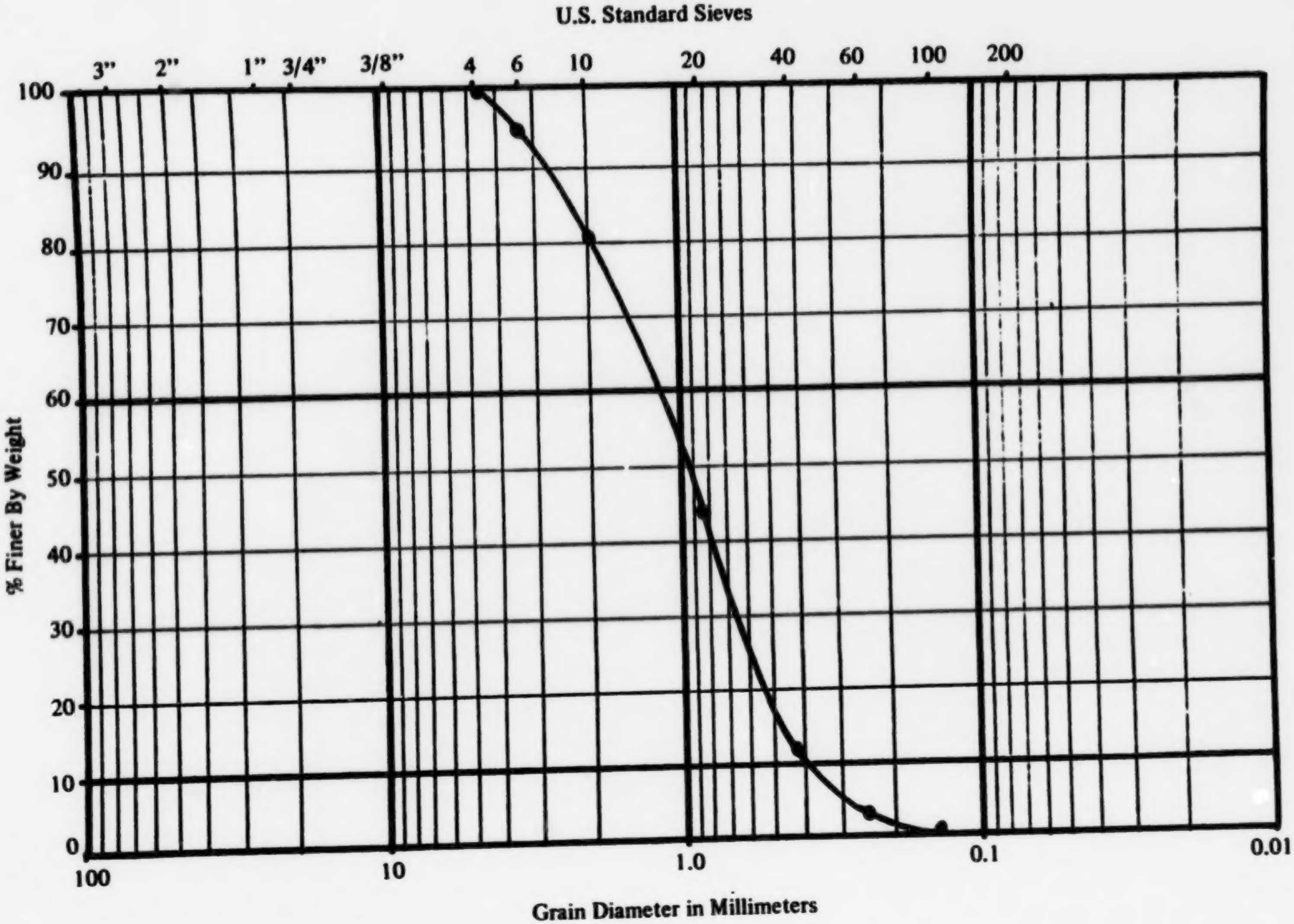


FIGURE 9-7
GRAIN SIZE CURVE

Such media are frequently used in experimental investigations, but most designers have not considered the extra cost justified in full scale installations. Important exceptions should be noted, however: One equipment supplier, Dravo, emphasizes use of uniform coarse media in deep beds; Baumann (7) recommends use of uniform anthracite and sand in dual media filters pointing out that the extra cost is probably not more than 1 percent of the overall cost of the filters.

9.5.2 Dual and Multi Media

Upflow washing stratifies a bed in accordance with the settling velocities of the media particles as determined by their size, shape and specific gravity (See Section 9.7). In a dual or tri-media bed, although each media component is still graded fine to coarse in a downward direction, lighter coarser media can be maintained above finer denser media. This makes it possible to approximate a coarse-to-fine gradation in down flow filtration units. Another advantage of dual or multi-media over a single medium is that mud balls formed in the filter remain above the coal-sand interface where they are subject to auxiliary scrubbing action (1).

The maximum settling velocities of media particles also determine the minimum wash rate required for adequate fluidization of the bed during backwash. Hence, for a given media size at the top of the bed, lower wash rates can be used if each media component is more uniform and the top portion of the filtration is of anthracite rather than a heavier material. Baumann and Cleasby (1) recommend that dual or multi-media be sized so the coarsest (d_{90}) sizes of each component have about the same minimum fluidization velocity.

9.5.3 Pore Size and Intermixing of Media Components

The hydraulic behavior and filtration performance of any given media are more properly related to pore size than to grain size. For single media component, pore size is directly proportional to grain size, and the porosity (percent of volume represented by pores) is a constant depending only on media shape. Coal which tends to be angular has a porosity of almost 0.5 whereas sand porosity is closer to 0.4. In water treatment applications, coal media, because of its greater porosity, has been found to give poorer removals but lower pressure losses than sand of the same grain size (45).

The pore size in multi-component filter media depends on the degree of intermixing of the components. With no mixing, pore size distribution simply follows that of the components. With intermixing, however, the finer layers of the denser material below are dispersed into the voids of the coarser layers of lighter material above. No precise methods have as yet been demonstrated for calculating actual pore size, or even the degree of intermixing, from the characteristics of media components. Where such information is of interest it may be obtained from test columns or from experience with specific combinations of components in other installations. Limiting size ratios have been proposed to control intermixing and to avoid the extreme where lower density coarse media is overtopped by very fine high density media. Camp (46) has hypothesized that for dual media filters with an interface size ratio (coarsest coal/finest sand) of 2.8 no intermixing should result, whereas for a ratio of 4.0

intermixing would occur over a depth of about 5 inches. Baumann (7) indicates only limited intermixing (6 inches) and no overtopping with highly uniform coal and sand having a size ratio of 3.35.

Culp and Conley (47) indicate that to avoid overtopping in dual media beds, the effective size of the coal grains must be no more than about three times the effective size of the sand. Whatever the exact limiting ratio, designs using coarser anthracite to accept higher solids loadings at lower head losses must also use correspondingly coarser sand. Where stringent effluent quality or weak floc conditions require a finer media component than the sand, garnet can be used. Coal/garnet size ratios as high as five will not result in overtopping.

The significance and desirability of intermixing in dual or multimedia beds is a subject of debate. Camp (46) has reported deliberate selection of dual media sizes to minimize intermixing, whereas some manufacturers actively promote intermixing as advantageous in three and four component media, claiming that controlled intermixing approximates a "theoretically ideal" coarse to fine gradation of voids in the direction of filtration.

In side by side tests at Washington D.C. (See Table 9-3) on chemically treated effluent, mixed (tri) media filters did show slightly better effluent quality performance than dual media filters. However there was no evidence to demonstrate that this was due to intermixing rather than just to the fine, high specific gravity garnet present at the bottom of the filters.

9.5.4 Specific Media Designs

Table 9-5 lists typical characteristics for several specific media configurations which have been used in normal design of wastewater filters. All provide initial filtration through coarse media either by upflow filtration or by downflow filtration using dual, tri or deep, uniform, media. Configurations using fine single media and hence requiring special cleaning provisions, are not included. Typical application conditions (floc strength, solids load) are shown for each design. The dual-media designs for the most part employ depths ranging from 30 to 36 inches. To allow for level variations due to uneven backwashing, sand depths are set at 12 to 15 inches even though only the top few inches significantly affect removals Minimum depths for the anthracite are 15 to 18 inches. Greater depths may be necessary where solids loads are heavy.

In tri-media designs the overall depths and the minimum depths of anthracite and of the combined finer media are in the same range as in dual media designs.

The single media configurations employ depths of 60 inches or more. In downflow filtration this great depth is intended to improve efficiency, while in upflow units it has an additional purpose of adding weight to restrain the bed from uplift due to differential pressures during operation. Where uplift exceeds the submerged weight of the media it will either fluidize the bed or lift it in a "piston" effect (small diameter filters).

TABLE 9-5

TYPICAL MEDIA DESIGNS FOR FILTERS

Media Design	Coal			Sand			Garnet			Typical Application Conditions	Reference
	Dia. mm	Depth in	Unif. Coeff.	Dia. mm	Depth in	Unif. Coeff.	Dia. mm	Depth in	Unif. Coeff.		
Single	–	–	–	1-2	60	1.2	–	–	–	A	13
Single	–	–		2-3	72	1.11	–	–	–	A	48
Dual	0.9	36	< 1.6	0.35	12	< 1.85	–	–	–	B	29
Dual	1.84	15	< 1.1	0.55	15	< 1.1	–	–	–	A	7
Tri	1.0-1.1	17	1.6-1.8	0.42-0.48	9	1.3-1.5	0.21-0.23	4	1.5-1.8	B	25
Tri	1.2-1.3	30	–	0.8-0.9	12	–	0.4-0.8	6	–	C	13

NOTES:
A = Heavy Loadings, Strong Floc.
B = Moderate Loadings, Weaker Floc.
C = Moderate Loadings, Strong Floc.

Some of the theoretical advantage of upflow coarse-to-fine filtration is lost because minimum grain sizes must be coarse enough to avoid excessive uplift.

Additional resistance to uplift is provided in many upflow designs by placing a restraining grid on top of the media. The spacing between bars of the grid must be large enough to prevent upward bed movement during filtration. Although these two requirements appear contradictory, arching of the grains takes place between the bars, allowing a reasonably large spacing, in the range of 100 to 150 times the diameter of the smallest grain size in the beds.

9.5.5 Selection of Media

Pilot testing is indispensable to provide the information necessary for meaningful comparison of different media designs or to assure the effluent quality performance of any media design selected. Without pilot testing, the designer should select a media which, on the basis of experience with similar influents, may be expected to provide good solids removal with low head loss buildup. In general, any such media would include an ample depth of coarse media followed by fine media in the size ranges indicated for dual media configurations in Table 9-5.

Pilot testing to guide media selection should define headloss development vs time for each media design, under all test conditions. Suggested ranges for test conditions are given in Section 9.9. If one media design clearly gives lower head buildup at all times and under all test conditions it may be selected directly provided its backwash requirements are not extraordinary. If different media provide essentially the same headloss development over the range of test conditions, selection may be based on other factors. Where different media appear significantly better under different conditions, selection should be based on cost comparison of the alternative designs each at its most favorable rate, terminal headloss and run length, determined as indicated in Section 9.4. Significant differences in backwash flows should be taken into account.

9.6 Filter Control Systems

Major filter functions requiring monitoring and or control are:

1. Head Loss
2. Effluent quality
3. Initiation of backwash where automatic
4. Flow rate through the filters
5. Backwash sequence, rate and duration.

An important tool in performance control is the automatic turbidimeter which can continuously monitor the filter feed and product. This allows the operator to anticipate difficulties from changes in feed quality, and rapidly remedy process failures. In addition, these devices allow the operator to rapidly evaluate the effects of changes in process variables and provide a continuous record of plant performance. All turbidimeters operate on the pinciple of measurement of scattered or transmitted light. A variety of commercial instruments are available.

Filter installations should be equipped with appropriate loss of head and flow indicators. Individual filters should have multiple taps for pressure readings if full scale experimental testing is desired.

Provisions for automatic or remote initiation of backwash by timers or based on headloss or turbidity monitoring may be justified to reduce the need for operating attention.

Three types of flow control systems are used for filters:

1. Effluent rate control
2. Influent flow splitting
3. Variable declining rate control.

Features of these systems and of automatic backwash systems are described below.

9.6.1 Effluent Rate Control

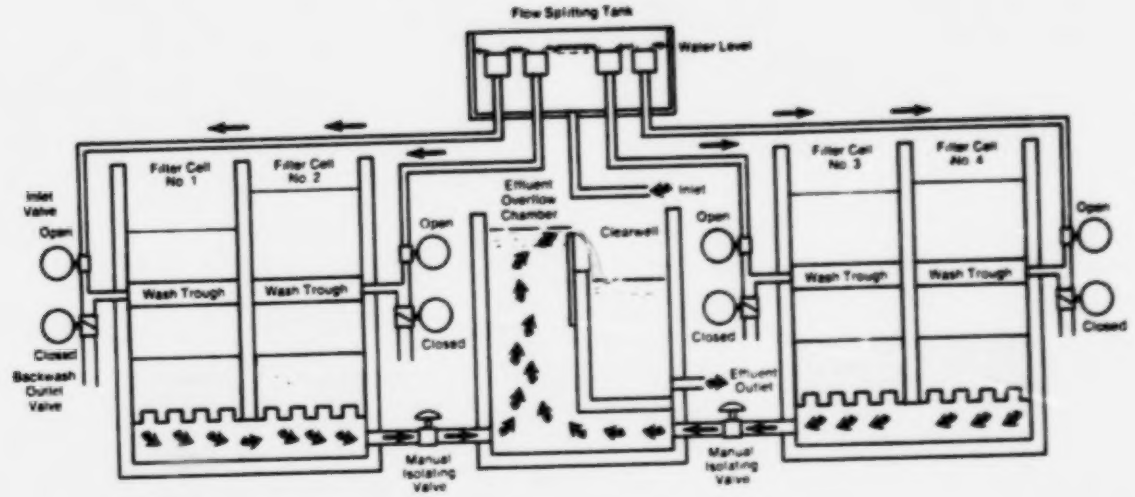
This system, common in traditional water treatment plant designs, maintains a set flow for each filter by throttling the effluent (see Figure 9-1). The throttle valve may be controlled directly by mechanical linkage to a venturi controller or indirectly by a set point controller linked to a pneumatic or hydraulic valve operator. The direct acting system is unsuitable if flows to individual filters must vary over the day. The indirect system is complex and both may be troublesome in maintenance. The system is also wasteful of head since available head not needed in a clean filter is lost in the controller. In addition, control valves may produce high frequency surges in the filter bed with accompanying loss of efficiency (10).

9.6.2 Influent Flow Splitting

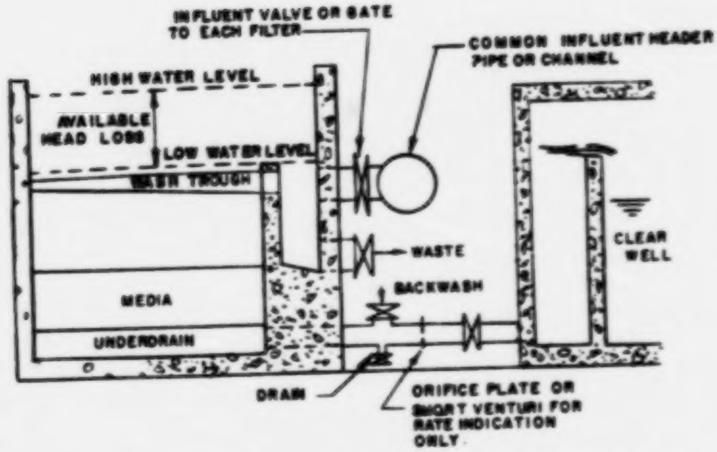
In this system flow is evenly divided among filters in a splitter box located at or above the level of the top of the filter boxes (see Figure 9-8a). The boxes themselves are made deep so that the water level in them can build up to provide the maximum operating head needed when the filter bed is dirty. A weir on the filter outlet maintains a constant back pressure or minimum water level to prevent accidental dewatering of the bed.

Advantages of influent flow splitting include:

1. Rate controllers with attendant maintenance and surging problems are eliminated.


Flow Splitting Tank
Water Level
Filter Cell No 1
Filter Cell No 2
Filter Cell No 3
Filter Cell No 4
Inlet Valve
Open
Closed
Backwash Outlet Valve
Wash Trough
Wash Trough
Effluent Overflow Chamber
Clearwell
Inlet
Open
Closed
Open
Closed
Effluent Outlet
Wash Trough
Wash Trough
Open
Closed
Manual Isolating Valve
Manual Isolating Valve

A. INFLUENT FLOW SPLITTING
(Courtesy of Permutit, Inc.)


INFLUENT VALVE OR GATE TO EACH FILTER
COMMON INFLUENT HEADER PIPE OR CHANNEL
HIGH WATER LEVEL
AVAILABLE HEAD LOSS
LOW WATER LEVEL
WASH TROUGH
WASTE
CLEAR WELL
BACKWASH
MEDIA
UNDERDRAIN
DRAIN
ORIFICE PLATE OR SHORT VENTURI FOR RATE INDICATION ONLY

B. VARIABLE DECLINING RATE FILTRATION (1)

FIGURE 9-8
FLOW CONTROL SYSTEMS

2. Flow variations are distributed to filters automatically.

3. Head loss may be read directly from water levels in filter boxes.

4. Only a single master flow meter is needed.

5. Changes in filter rate are gradual because of time required for head to build up in filter boxes.

Disadvantages are:

1. The head not needed for filter operation is lost in the drop between the splitter and the filters.

2. Capital cost of filter box construction is increased by the greater depth.

9.6.3 Declining Rate Filtration

This system requires multiple filters. All operate under the same head but at different flow rates depending on the degree of clogging. Under constant head the output from a single filter declines as the run progresses. The filter selected to be backwashed is always the one which has been on line the longest and is most clogged. Total output from all filters is controlled by varying the head applied. Figure 9-8b shows a variable declining rate filter.

The head on the filters may be controlled by varying either the upstream or downstream water level (10) (49). With downstream water level control, an equalizing chamber must be provided to limit the rate of change of head and hence of flow, when filters are taken off line or restored to service. It is common to apply maximum design loadings to the filters as a group and to limit maximum rates on individual clean filters to from 20 percent to 40 percent above these design loadings.

Advantages cited for declining rate filtration (10) (49) include better effluent quality, absence of surges, and significantly lower total head requirements. Less head is needed because:

1. There is no loss due to throttling or due to free fall after flow splitting.

2. As rate declines turbulent head losses (underdrains, valves, etc.) reduce rapidly (in proportion to second power of flow) making head available to overcome resistance of clogged filter (proportional to flow).

For proper operating control, flow rates should be measured individually for each declining rate filter. Only single indicators are needed for inlet and outlet levels on head loss, since these are the same for all filters.

The chief disadvantage of this method of flow control is the need for a large volume of water storage upstream of the filter.

9.6.4 Backwash Control

Programmed backwash systems are widely used in current designs. Such systems consist of interlocked controllers and timers programmed to open and close valves, make or break siphons, start and stop pumps and blowers and limit backwash flows to control the rate, duration and sequence of activities during backwash. Even where backwash is manually initiated, the rest of the control system may be entirely automatic.

Proprietary systems, with various features are available from different manufacturers. One such system, designed to operate with only a single control valve is shown in Figure 9-9.

9.7 Filter Cleaning Systems

9.7.1 Upflow Washing

Accumulated solids are removed from filters by a rapid upflow of washwater. The waste flow is then recycled to some prior treatment unit, usually primary settling. Washwater sources may include filter influent, filter effluent or effluent from subsequent treatment units. Storage of washwater supply may be needed if rates required exceed the flow available. Recycled spent washwater flows should be equalized by storage so they do not disrupt prior treatment processes. Backwash rates for most effective cleaning vary with media size and density.

Baumann and Cleasby (1) recommend providing upflow wash capacity adequate to fluidize the 90 percent finer size of each media component at the warmest expected water temperature. Figures 9-10 and 9-11 may be used to determine minimum upflow velocities or wash rates to fluidize coal, sand, and garnet media of various sizes. Rates should be variable to compensate for changes in temperature, viscosity, and hence bed expansion. Maximum hydrodynamic shear and most efficient cleaning have been shown to occur when the porosity of the expanded media is about 0.7 (1). To reach this porosity in the surface layers of a non-uniform sand (effective size = 0.4, uniformity coefficient = 1.47) requires almost 50 percent expansion (7). Because of its higher unexpanded porosity coal requires only 20 to 25 percent expansion to reach a porosity of 0.7 in the surface layers. In practice the backwash rates generally used in filter designs, range up to 20 gpm sq ft, and do not provide more than 15 to 30 percent expansion except for very fine media (50). This means that higher backwash durations (5 to 10 or at the extreme 15 minutes) and somewhat higher washwater consumption are required than at rates which would provide the most efficient washing.

Whatever wash rate and duration are expected, design of piping, valves, pumps and storage tanks should provide extra capacity of at least 25 percent.

Methods are available for predicting expansion of sand beds accurately (51), but have not yet been satisfactorily extended to other media components. Expansion of multi-component media can be rapidly obtained, however, from backwash tests in pilot columns.

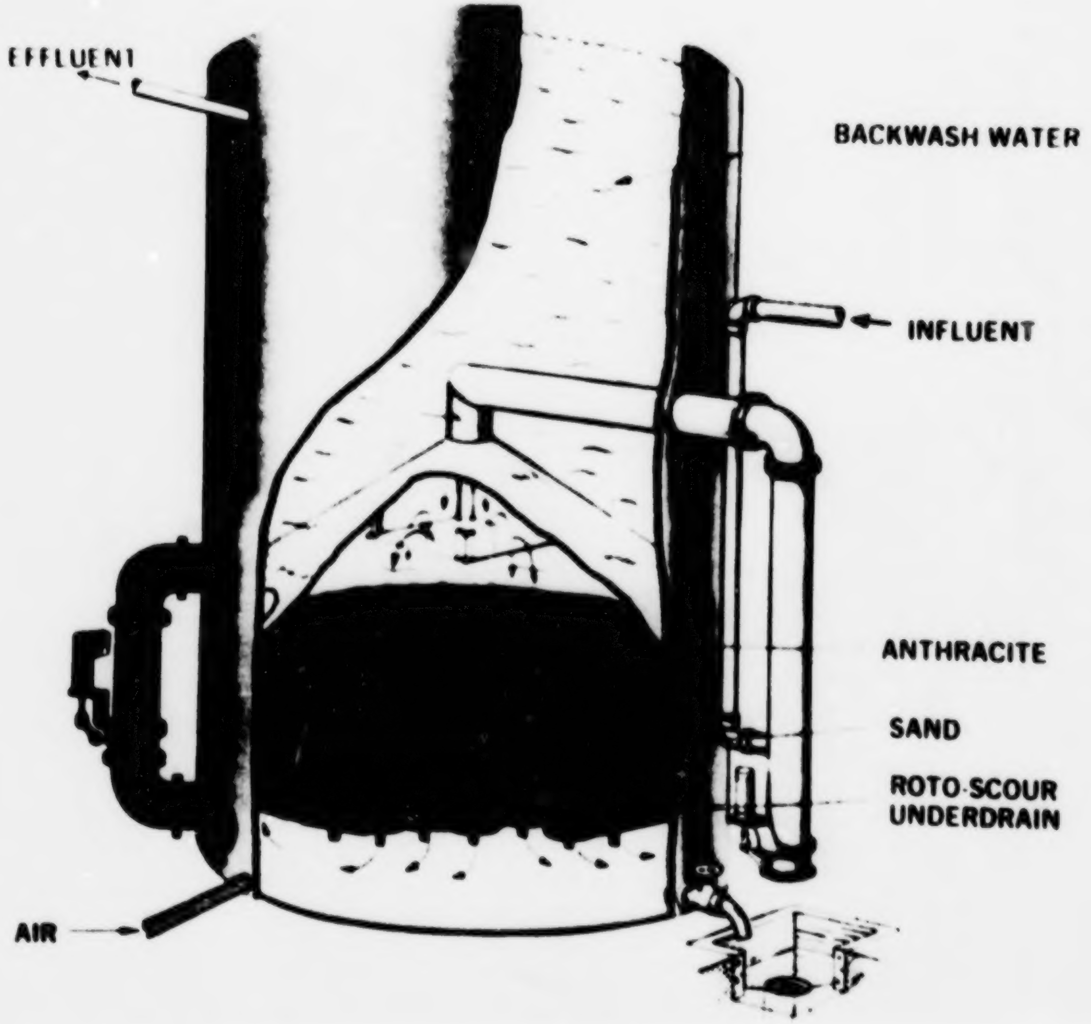

EFFLUENT
BACKWASH WATER
INFLUENT
ANTHRACITE
SAND
ROTO-SCOUR UNDERDRAIN
AIR

FIGURE 9-9
AUTOMATIC GRAVITY FILTER, SINGLE COMPARTMENT
(Courtesy of Ecodyne Corp.)

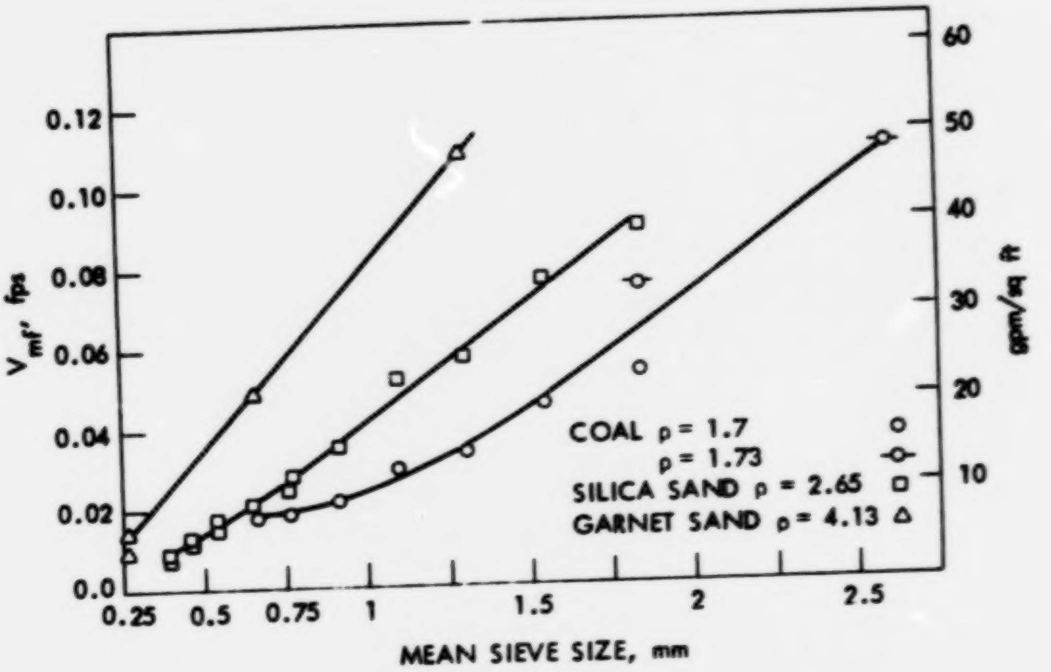


FIGURE 9-10

MINIMUM FLUIDIZATION VELOCITY, Vmf,
TO ACHIEVE 10 PERCENT BED EXPANSION
AT 25°C (1).

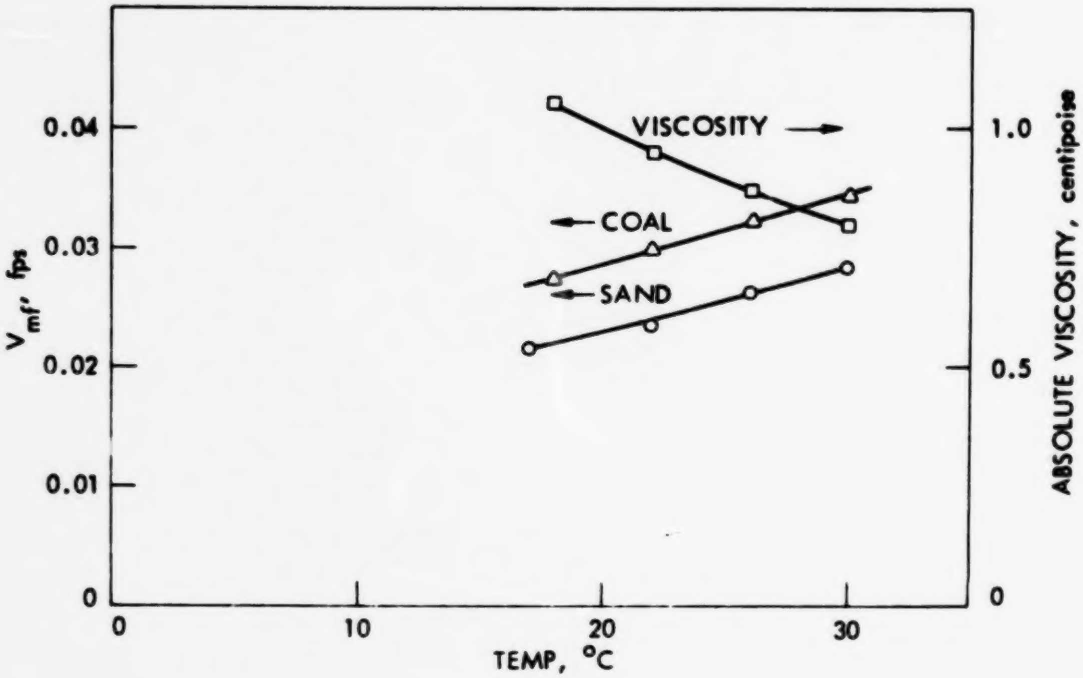


FIGURE 9-11
EFFECT OF TEMPERATURE ON Vmf FOR
SAND AND COAL AND ON ABSOLUTE VISCOSITY
OF WATER (1)

Several manufacturers offer coarse media filter systems with wash rates too low to fully fluidize the bed. Auxiliary air scouring is provided, but its long term effectiveness in maintaining the media in good condition has not been proven. Studies at Iowa State University (52) indicate that air scour and simultaneous upflow washing at full fluidization can free a dirty filter bed of mud balls which accumulate during multiple cycles using air scour followed by washing. Further work will study effectiveness of air scour and simultaneous upflow washing at less than full fluidization.

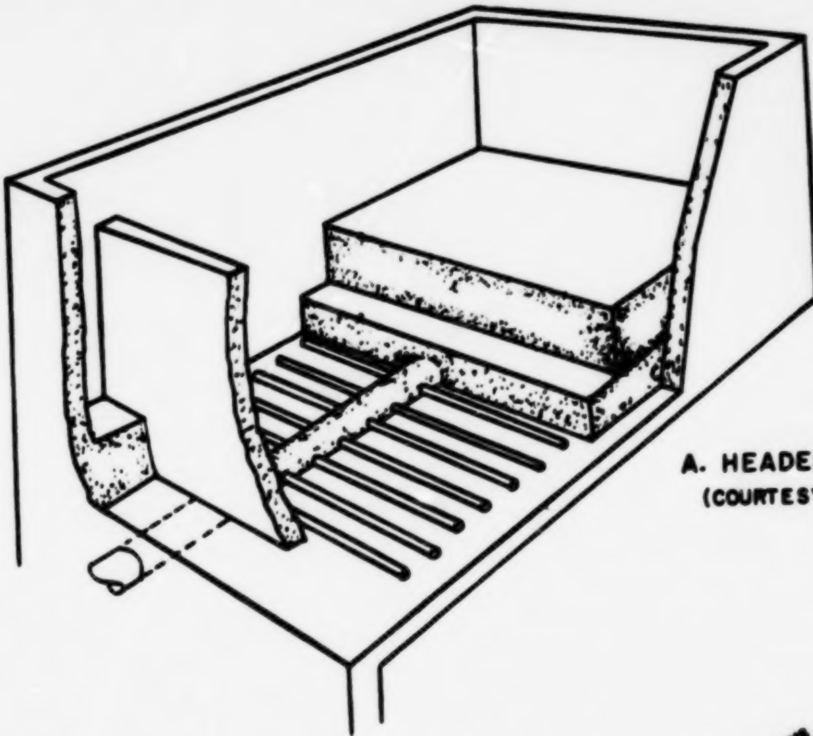
9.7.2 Underdrains

Underdrains should distribute washwater as uniformly as possible over the area of the filter. Excessive variation in washwater rate results in uneven and ineffective cleaning. Moreover, the accompanying excessive jet action can lead to lateral displacement of gravel and clogging of the underdrains with filter media.

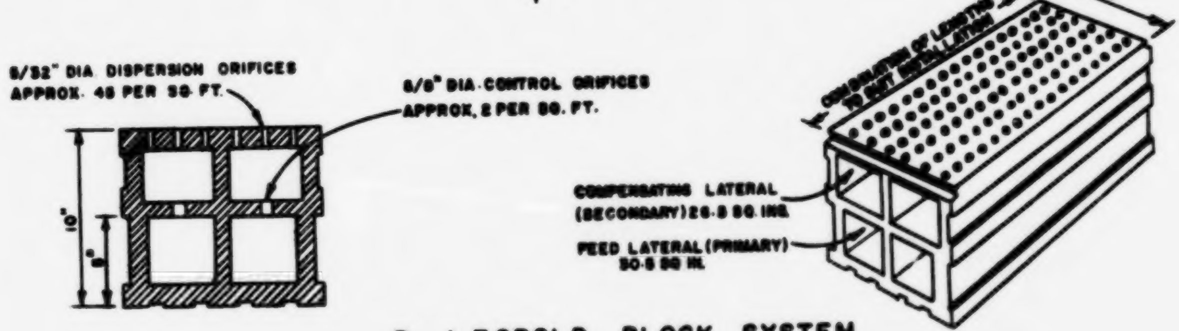
In general, underdrain systems developed for water filtration may also be used in wastewater applications. One of the first systems employed in water filtration consists of several layers of graded gravel surrounding manifold piping positioned on the filter floor. Orifices in the manifold piping provide preliminary distribution of the washwater as shown in Figure 9-12a. The final distribution is accomplished as the water moves upward through the gravel. Fair and Geyer (44) provide rules of thumb for sizing lateral header systems and a basis for hydraulic analysis.

Several commercial systems are available which employ patented false bottom distributors in place of the manifold. The gravel is then placed on top of the false bottom. Leopold Block andWheeler Filter Bottom systems are illustrated in Figure 9-12b and c. A system manufactured by Dravo (Figure 9-12e) distributes both air during airwash and washwater during normal backwash.

Table 9-6 shows the standard gravel design used in water filtration together with a modified design suggested by Baylis (45) for use with higher backwash rates.



A. HEADER LATERALS
(COURTESY OF THE AWWA)


5/32" DIA DISPERSION ORIFICES
APPROX. 48 PER SQ. FT.
5/8" DIA. CONTROL ORIFICES
APPROX. 2 PER SQ. FT.
10"
6"
COMBINATION OF LENGTHS TO SUIT INSTALLATION
11"
COMPENSATING LATERAL
(SECONDARY) 26.8 SQ. INS.
FEED LATERAL (PRIMARY)
30.5 SQ IN.

B. LEOPOLD BLOCK SYSTEM

(Courtesy F. B. Leopold Co.)

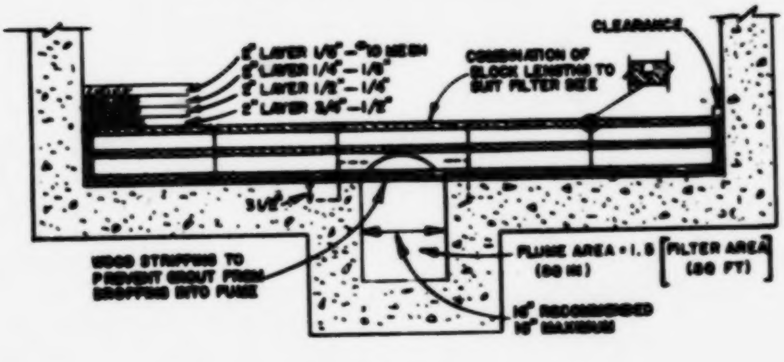
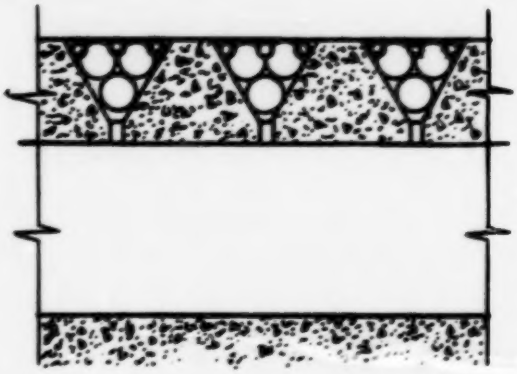
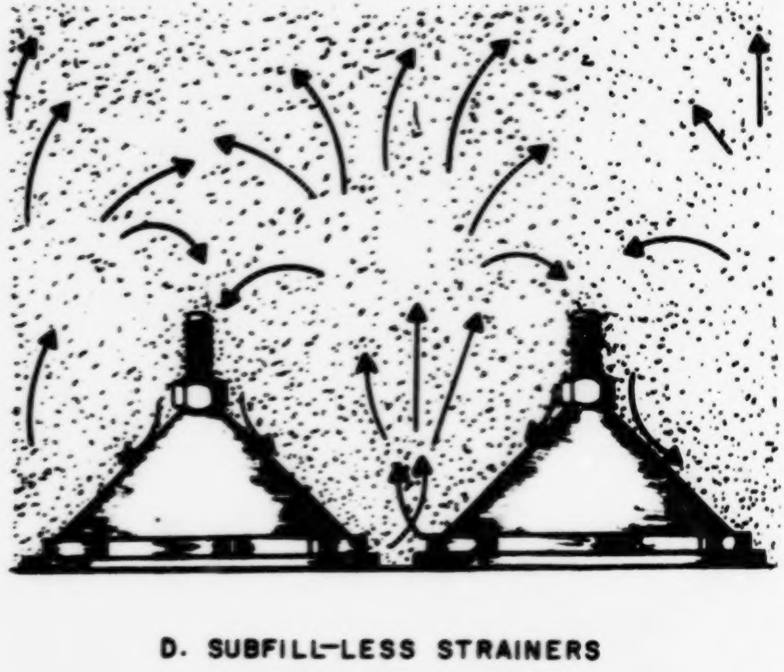

2" LAYER 1/8" – #10 MESH
2" LAYER 1/4" – 1/8"
2" LAYER 1/2" – 1/4"
2" LAYER 3/4" – 1/2"
COMBINATION OF
BLOCK LENGTHS TO
SUIT FILTER SIZE
CLEARANCE
3 1/2"
WOOD STRIPPING TO
PREVENT GROUT FROM
DROPPING INTO FLUME
FLUME AREA = 1.5 [FILTER AREA (SQ FT)]
(SQ IN)
16" RECOMMENDED
18" MAXIMUM

FIGURE 9-12

UNDERDRAINS

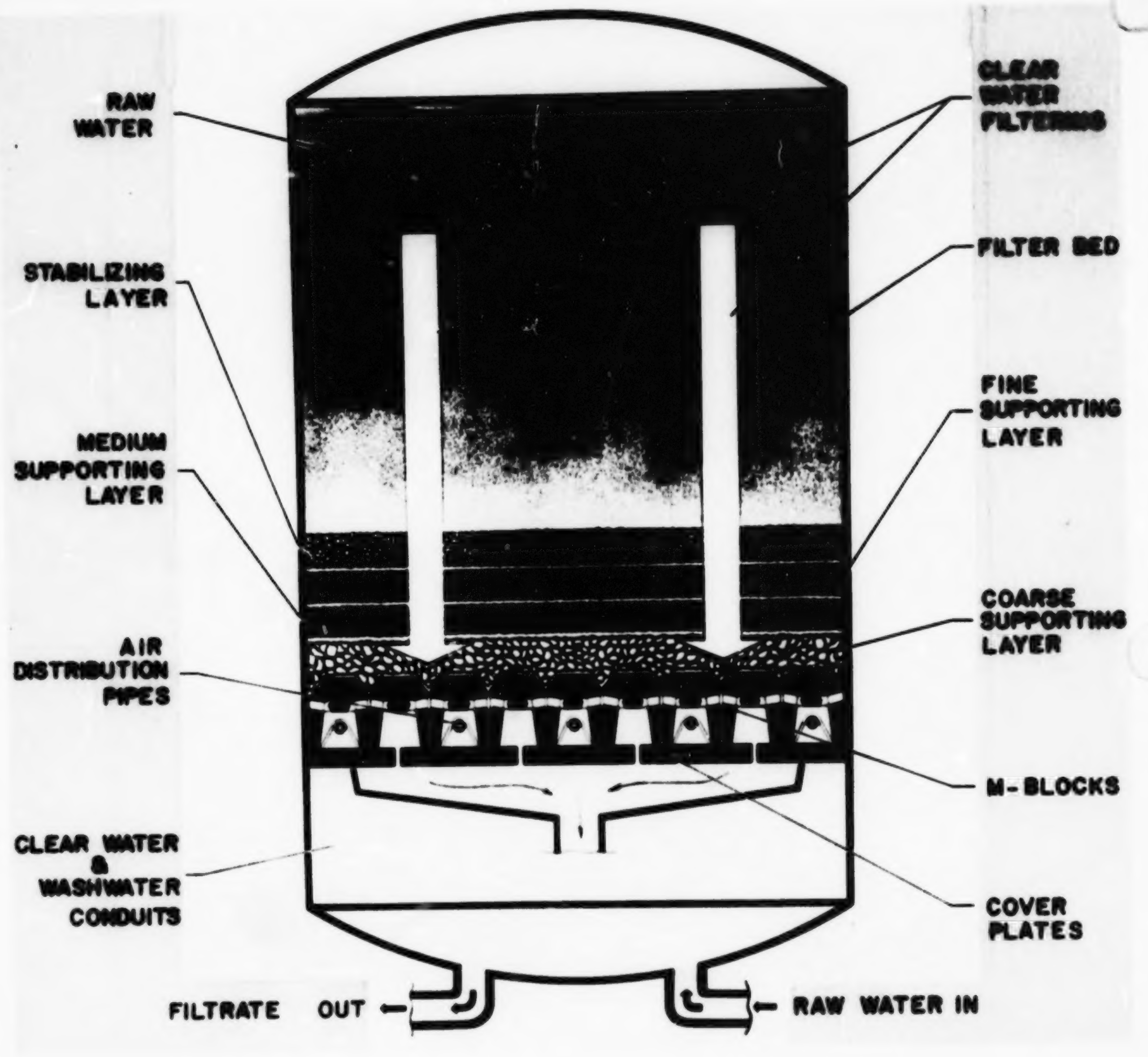


C. WHEELER FILTER BOTTOM
(Courtesy of B I F)



D. SUBFILL–LESS STRAINERS
(Courtesy of Ecodyne Corp.)

FIGURE 9-12 (Continued)


RAW WATER
STABILIZING LAYER
MEDIUM SUPPORTING LAYER
AIR DISTRIBUTION PIPES
CLEAR WATER & WASHWATER CONDUITS
FILTRATE OUT
CLEAR WATER FILTERING
FILTER BED
FINE SUPPORTING LAYER
COARSE SUPPORTING LAYER
M-BLOCKS
COVER PLATES
RAW WATER IN

E. DRAVO M-BLOCK SYSTEM
(COURTESY OF DRAVO CORP.)

FIGURE 9-12 (continued)

TABLE 9-6

FILTER GRAVEL DESIGN

Standard		Baylis	
Depth in.	**Size** in.	**Depth** in.	**Size** in.
2-1/2	1/12 to 1/8	5	1 to 2
3-1/2	1/8 to 1/4	2	1/2 to 1
3-1/2	1/4 to 1/2	2	1/4 to 1/2
2	1/2 to 3/4	4	1/8 to 1/4
4	3/4 to 1-1/2	2	1/4 to 1/2
6	1-1/2 to 3-1/2	2	1/2 to 1
		4	1 to 2

Baylis found that with the standard design the upper layers of gravel could fluidize under high backwash rates (above 20 gpm/sq ft), and proposed placing a final layer of heavy gravel over the finer gravel to prevent this fluidization. A 3-inch layer of coarse (1 mm) garnet or ilmenite above the gravel has also been suggested to overcome this problem (8).

Some systems eliminate the gravel layers entirely by using nozzles or porous plates. Advantages and disadvantages of these systems are discussed elsewhere (1) (8). One system, manufactured by Graver Water Conditioning Co., utilizes a cast concrete false bottom with plastic or metal strainers on about 12-inch centers as shown in Figure 9-12d.

9.7.3 Washwater Troughs

Except for small units, filters are commonly equipped with washwater troughs spaced on about 6-ft centers. Hydraulic design of these devices is discussed in Fair and Geyer (44).

Troughs are installed with the aim of aiding uniform distribution of washwater and avoiding dead spots which may retard the removal of dislodged solids from the filter box.

9.7.4 Auxiliary Cleaning

The function of auxiliary cleaning by air scour or surface wash is to loosen accumulated deposits from the filter media. The slimes and organic particulates encountered in wastewater filtration cannot be completely loosened by normal backwash flow.

Typical surface wash equipment consists of either fixed or rotating pipe distributors fitted with nozzles which are placed about 1 to 2 inches above the top of the bed. While the surface wash is on, the backwash expansion is set at a lower rate than after the surface wash is terminated. Surface washwater is supplied at 50 to 100 psi at rates approximating 1 to 3 gal/min/sq ft of bed.

With the widespread use of media permitting deeper floc penetration, the ability of the above type of surface wash to clean mid-and lower portions of the bed has been questioned. Wash jets positioned at lower levels in the bed may help to alleviate this problem.

Air scour systems have been increasingly used in an attempt to reduce washwater requirements and to effect cleaning of the deeper portions of the bed. Some concern has been expressed concerning loss of finer lighter media particles when air washing is used (8). Where this is a problem, air scour should be applied separately from the backwash, with liquid in the filter box drawn down below the washwater overflow level so that no overflow occurs during air wash. Allowance must be made for 6 to 9 inches of water level rise due to air lift (7). Although some of the lighter media may remain on the surface of the water and subsequently be lost, the rate of such loss should generally be negligible.

To prevent air scour from disrupting gravel placement, air is usually injected through a grid above the under-drain gravel. It may go directly into the underdrains where no gravel is used.

9.7.5 Backwash Sequence

The cleaning cycle time (total downtime during one cleaning operation) includes time for valve openings and closings, time for drainage of inflow from the filter and time for the actual upflow washing and auxiliary cleaning. Unless influent in the filter box is to be wasted, drainage time should be calculated at normal filter rates. Valve openings and closings to start and stop backwash and air scour should be gradual to keep from upsetting the media gradation and structure.

A typical sequence for backwash with auxiliary surface wash is:

1. Shut influent and permit water level to drain down to top of the washwater troughs or other washwater control weir.

2. Apply surface wash for 1 to 3 minutes.

3. Apply upflow wash and surface wash together 5 to 10 minutes as needed to flush out solids.

4. Shut off auxiliary wash and apply backwash alone for 1 to 2 minutes at rate needed to classify the bed.

5. Return bed to service.

A typical sequence for cleaning using upflow wash and air scour is:

1. Stop influent and lower the water level to a few inches above bed.

2. Apply air alone at 2 to 5 cfm/sq ft for 3 to 10 minutes.

3. Apply water backwash at 2 to 5 gpm/sq ft with air on until water is within one foot of wash water trough.
4. Shut air off.
5. Continue water backwash at normal rate for usual period of time.
6. Apply backwash for 1 to 2 minutes at a rate required to insure hydraulic classification of the filter media.
7. Return bed to service.

9.8 Filter Structures and General Arrangement

A typical wastewater filter consists of a tank or filter box containing an underdrain system, media and sufficient overall depth to contain media during backwash. In gravity units, the overall depth must also provide for operating submergence and freeboard. Influent, effluent, washwater and waste connections are provided. In addition, all wastewater filters should have provisions for auxiliary cleaning.

Underdrains are designed to properly distribute the washwater during cleaning (see Section 9.7). During normal operation, underdrains collect filter effluent (downflow operation) or distribute influent (upflow operation). Washwater troughs and filter inlets (downflow) or effluent launders (upflow) are located in the submerged zone above the media.

The recommendations for general arrangement and special structural features of concrete filters presented in Water Treatment Plant Design (53) are fully applicable to wastewater treatment applications.

For gravity filters of concrete construction, filter boxes are usually arranged in rows along one or two sides of a common pipe gallery, narrow side toward the gallery. This maximizes common wall construction and minimizes piping runs. Gravity filters may be of concrete or steel shell construction. Concrete units are generally square or rectangular and steel units circular. Sizes of gravity concrete units are limited to about 1000 sq ft (8); steel units are generally smaller.

For filters using influent flow splitting (see Section 9.6) multiple filter boxes have been constructed as compartments in a single round or square tank (concrete or steel) with common influent and waste piping located above the center of the tank and common washwater and effluent piping around the outside base.

Steel shell package pressure filters are cylindrical units with either horizontal or vertical axes. To minimize piping runs horizontal units are usually placed in rows with common piping along the ends. Vertical units are arranged in either rows or clusters. Horizontal pressure units are less restricted in size than the vertical pressure units and hence are normally used for plant capacities above 1 to 1.5 mgd (8). Where pressure units are used it is essential that manholes be provided for interior access both above the bed and below the underdrains. Pressure filters should also be provided with a means for hydraulic removal of all the filter media, and with sight glasses for observation of the bed.

In municipal water filtration plant designs it is common to totally enclose the pipe gallery and to locate controls in an enclosed superstructure above the gallery and overlooking the filters. In northern climates the filters themselves are usually included under the superstructure.

Wastewater effluent temperatures are generally somewhat higher than the local natural waters, so in a given locality there is less justification for housing wastewater filters than water filters. Piping and valves need to be protected in climates where freezing occurs either by housing or by insulation and heating. Controls need housing or weather protected enclosures in any climate. Local controls for each filter should be placed in a location from which the backwashing filter can be observed.

9.9 Pilot Studies

Specific pilot study objectives may include:

1. Comparing performance of different media designs in a given application.
2. Establishing relations between flow rate, headloss and run length for a particular media and application.
3. Establishing limiting headlosses and rates to assure required effluent quality and no deterioration due to breakthrough.
4. Characterizing wastewater variability in terms of performance variations with a specific media.
5. Studying effects of variations in chemical coagulation, flocculation, biological processes or other pretreatment.

In all cases the variation of influent and effluent quality and headloss buildup with time should be noted. Studies aimed at media selection may be limited to variations of size and depth for a particular configuration or may involve parallel operation of quite different media configurations.

Media selection tests should cover the range of operating conditions which may occur in the actual design: typically flow rates of at least 6 to 8 gpm/sq ft.,headlosses to 30 ft (unless a lower limit is imposed by constraints) and runs to at least 24-hr (if headloss or breakthrough do not limit). Effects of influent quality variations are revealed by conducting long-term tests on the same influent. It is important however to include influent quality conditions representative of the entire range expected in operation.

Studies to select filter rates and terminal headlosses for a particular media will generally include more rates and cover a somewhat wider range than a study involving media selection. Also extra effort should be given to including *critical* influent conditions (e.g. solids loadings and floc strength).

Studies aimed at characterizing the filterability of a waste (see Section 9.1) would generally be run at a low rate (2 to 4 gpm sq ft) using a media expected to give high effluent quality but without excessive surface filtration. Changes in quality and headloss with elevation in the bed should be observed. The widest possible range of influent conditions should be included.

Studies to test pretreatment generally would employ a specific media and a single filter rate, but would attempt to vary influent conditions by adjustment of pretreatment.

In judging what are critical influent conditions, variations in operating records of existing pretreatment facilities should be thoroughly studied. Where facilities do not exist, records for similar treatment at other locations should be considered.

Where existing biological pretreatment is inconsistent, with frequent periods of upset, it would be prudent to determine and eliminate the causes of these upsets rather than count on filtration as a cure-all for resulting problems (33).

Standard methods for conducting filtration pilot studies have not been established. The following is a list of typical equipment and practices.

1. Multiple filter tubes of transparent material with a minimum diameter of 4 to 6 in. are utilized.

2. The tubes are fitted for either gravity or pressure operation.

3. A false bottom underdrain is utilized with either a gravel covered plate or strainer backwash system.

4. Flow to filter units is set by a combination of positive displacement metering pumps, weirs, control valves, etc. Declining rate control or influent flow splitting may be used where important to test for design.

5. Sample taps are provided above and below the media, as well as at other locations within the bed. Tap locations are generally located near the top of each type of media used. If the effects of media depth variations are to be studied, however, taps should be located at 3 to 4 in. intervals down the column. As an alternative, parallel multiple columns of different lengths may be used.

Additional details of pilot filters are given in various references on filtration studies (7)(54)(55).

Pilot studies as outlined above cannot adequately determine effects of cleaning system design parameters. Cleaning of the filter bed is difficult to simulate in pilot scale because of the small surface area of the beds utilized. The small area makes it impossible to study surface wash and air scour. Results of water backwash may not be representative because of the wall effect.

Information on cleaning performance can be obtained only in long term studies using large pilot installations with filtering areas of 4 sq ft or more. Because such studies are expensive, it may be desirable to design cleaning systems based upon experience from other studies (50) (51) (52).

9.10 Special Designs

9.10.1 Radial Flow Filters

Some recently developed filters employ horizontal radial flow through media contained in the anular space between concentric vertical cylinders. The inner cylinder acts as distributor and draw off points are distributed around the periphery of the outer cylinder which forms the filter vessel.

In the Simater unit (Figure 9-13) developed in England and marketed in the U.S. by Dravo, the sand media is continuously moved downward, drawn off, washed in a separate tank and returned to the top of the bed. A filter developed by Hydromation Corporation (Figure 9-14) has a batch external wash to clean its synthetic resin media.

A Simater filter was tested on biologically treated wastewater at the Essex River Authority in England. The unit was run in parallel with two upflow units (23). Media size for the Simater filter was 0.5 to 1.0 mm, comparable to one of the upflow filters but much finer than the other. Rates were not stated for the radial flow unit but were apparently in the same range as for the upflow units 4 to 6 gpm sq ft. All the filters gave SS reductions of 60 to 80 percent and effluent SS concentrations below 10 mg l.

The Simater showed marked tolerance for short slugs at high influent solids. Prolonged operation at high concentration resulted in clogging of the outlet screens, but this could be prevented by using higher rates of media washing.

9.10.2 Travelling Backwash Filter

Hardinge Corporation furnishes a fine media (0.48 mm sand), multi-compartmented filter in which each compartment (8-in width) can be backwashed without stopping filtration in the remainder of the filter. (See Figure 9-15). Backwash rates are similar to those for other filters (up to 15 gpm sq ft), but the backwash time for individual compartments is as low as 45 seconds. Hence, each compartment can be backwashed every 1 to 2 hours without excessive washwater use.

Each compartment has its own underdrain section. Media is supported on 1-in. porous

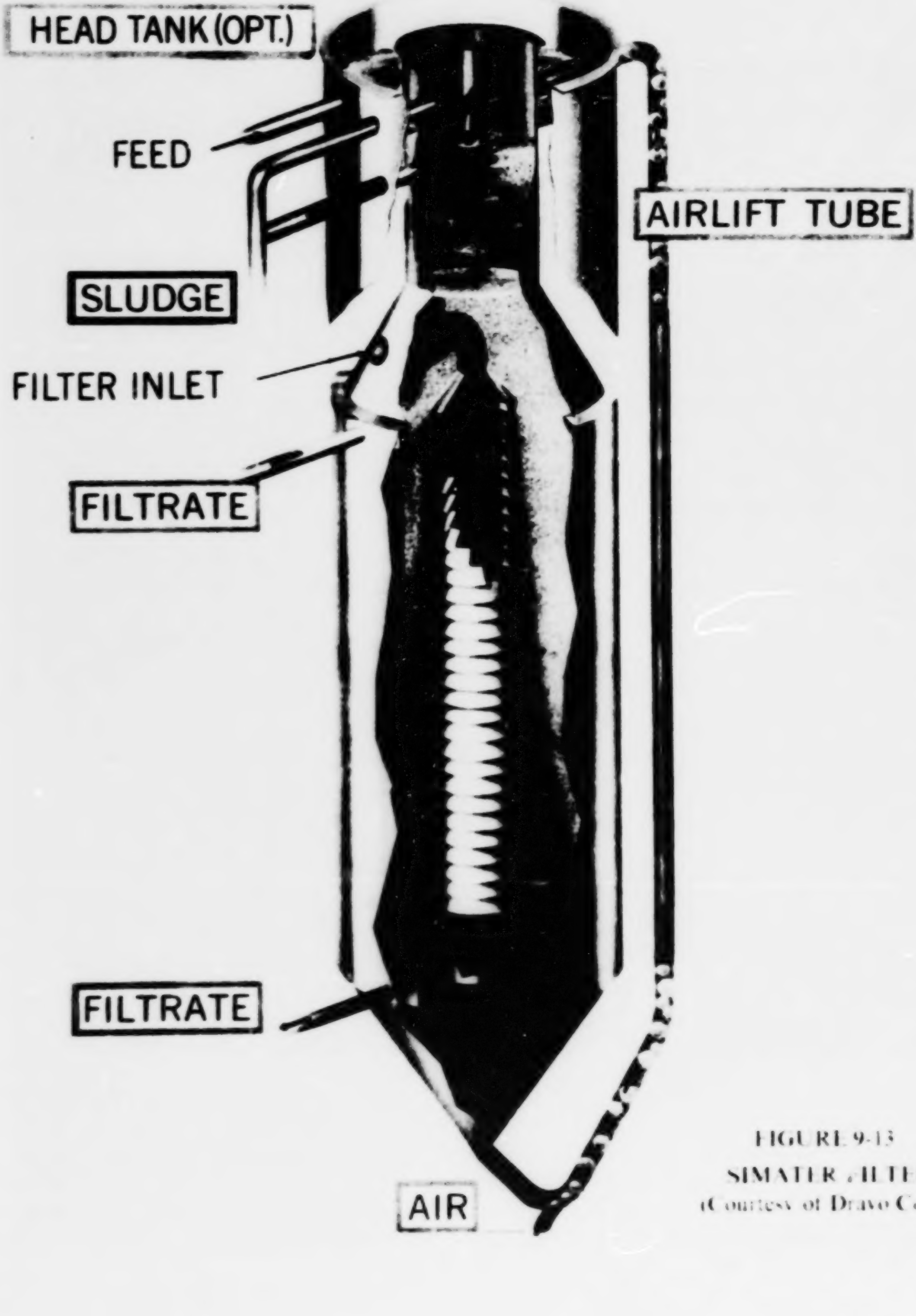

HEAD TANK (OPT.)
FEED
AIRLIFT TUBE
SLUDGE
FILTER INLET
FILTRATE
FILTRATE
AIR

FIGURE 9-13
SIMATER FILTER
(Courtesy of Dravo Corp.)

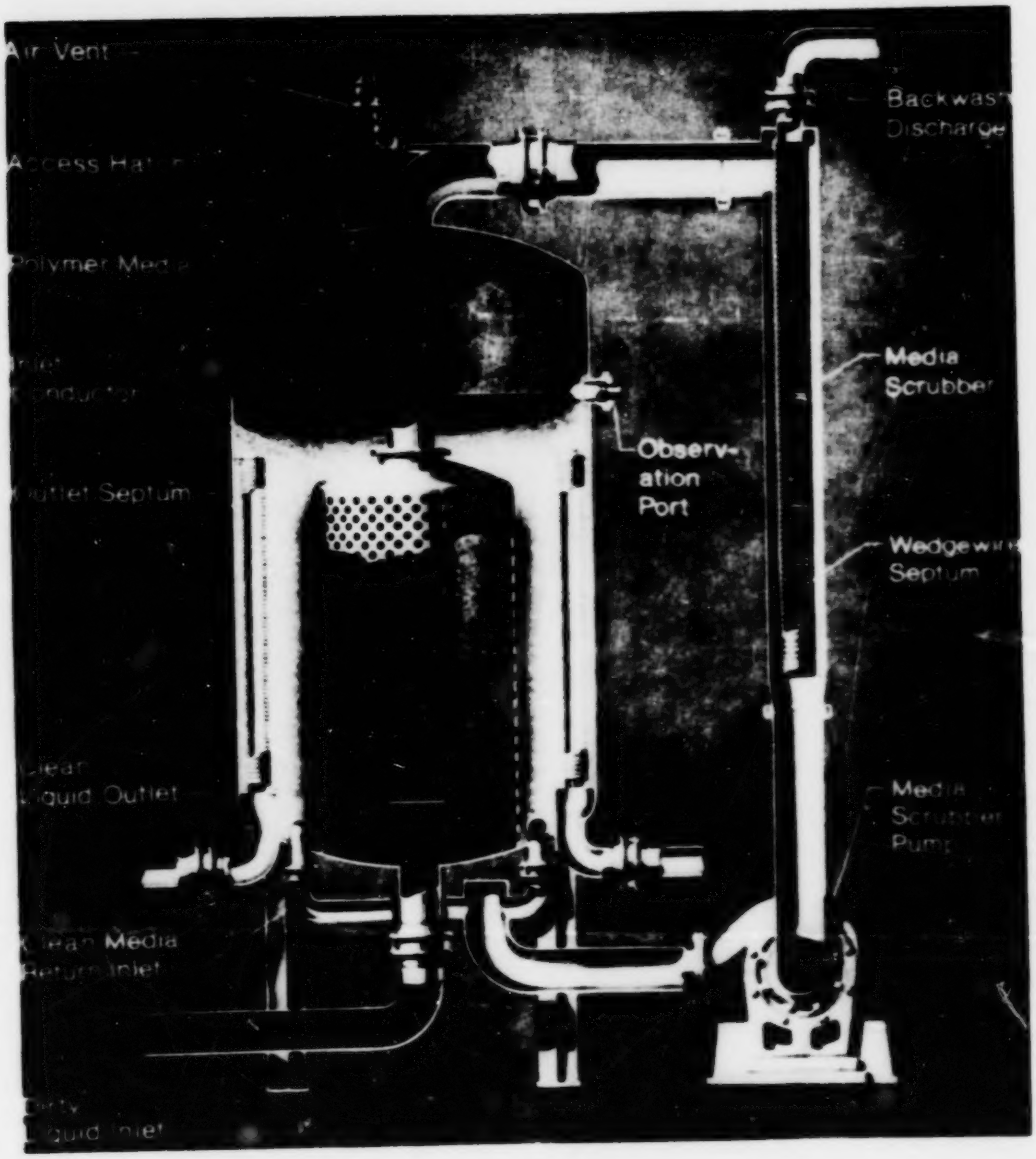
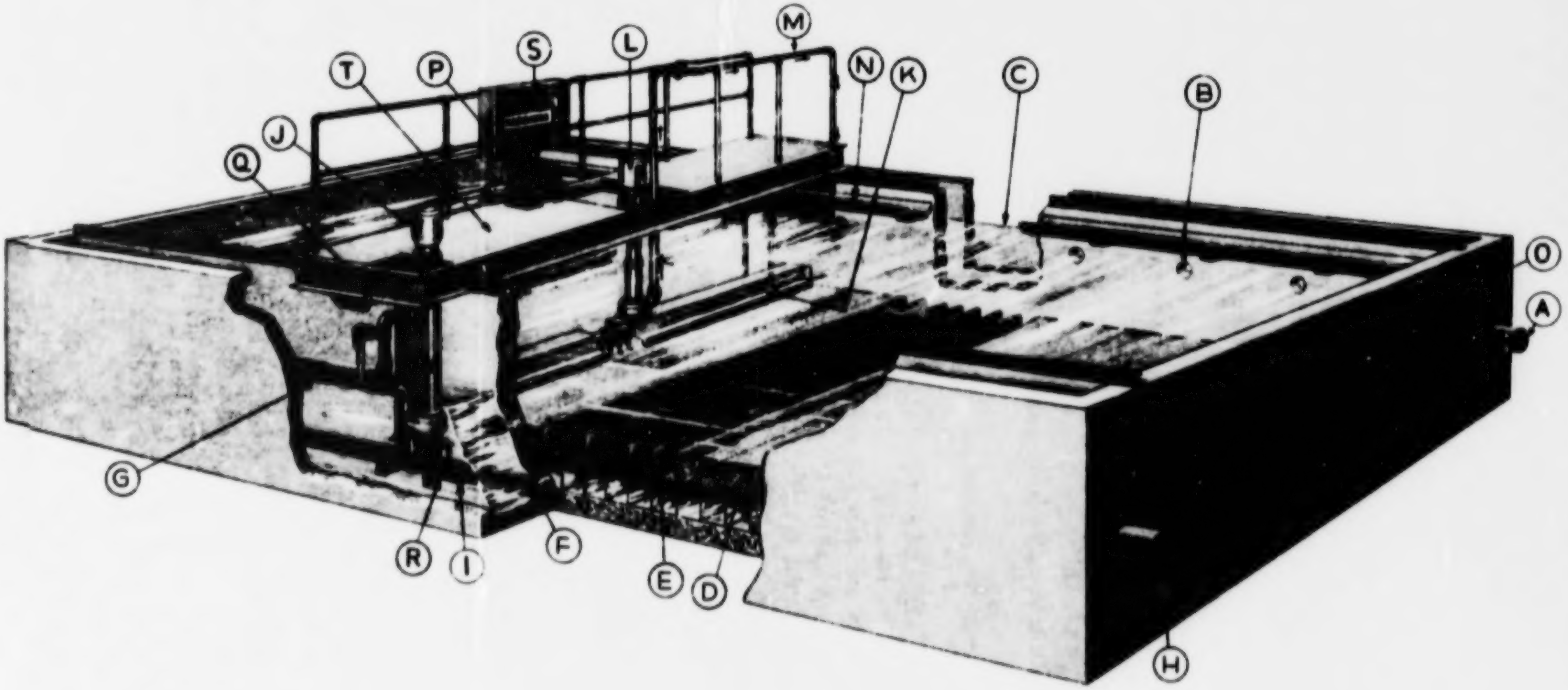

Air Vent
Access Hatch
Polymer Media
Inlet Conductor
Outlet Septum
Clean Liquid Outlet
Clean Media Return Inlet
Dirty Liquid Inlet
Backwash Discharge
Media Scrubber
Observation Port
Wedgewire Septum
Media Scrubber Pump

FIGURE 9-14
HYDROMATION IN-DEPTH FILTER



A. Influent line
B. Influent ports
C. Influent channel
D. Compartmented filter bed
E. Sectionalized under drain
F. Effluent and backwash ports
G. Effluent channel
H. Effluent discharge line
I. Backwash valve
J. Backwash pump assembly
K. Washwater hood
L. Washwater pump assembly.
M. Washwater discharge pipe
N. Washwater trough
O. Washwater discharge
P. Mechanism drive motor
Q. Backwash support retaining springs
R. Pressure control springs
S. Control instrumentation
T. Traveling backwash mechanism

FIGURE 9-15

HARDINGE AUTOMATIC BACKWASH FILTER
(Courtesy Koppers Co., Inc.)

plates over the underdrain. Flow from the underdrain sections discharges through individual ports to a common effluent channel.

The travelling backwash consists of a rolling bridge carrying two pumps and equipped with a hood extending over the length of a single compartment. The backwash pump draws water from the effluent chamber and discharges it into the underdrain section for the compartment where the bridge is stationed. The wash water pump withdraws backwash flow from the hood positioned over the compartment and discharges it to waste. Initiation of a backwash cycle is controlled either by timer or by headloss sensors.

Lynam (56) reported 68 percent removal of SS in uncoagulated activated sludge effluent by Hardinge filters at an SS loading of 0.5 lb/sq ft/day and 11.5 inches headloss. At 4.4 inches of headloss the removal at 0.4 lb/sq ft/day was 75 percent. At 11.5 in. headloss the maximum hydraulic loading was 6.0 gpm/sq ft compared with 2.5 gpm/sq ft at 4.4 in. In the same study, coagulation with alum did not improve performance.

9.10.3 Filter with On-Line Surface Scouring

Hydro-Clear Corporation offers a fully automatic, shallow bed, fine-media sand filter for tertiary wastewater treatment. The media consists of 10 inches of 0.45 mm sand with a uniformity coefficient of 1.5 supported on a wire mesh above the underdrain system. This filter combines air mixing of the water above the bed with air surging upward through the bed to prolong run length. Influent flow splitting controls the flow to parallel units.

Typically, a filter run consists of a preset number of filter cycles. Each cycle begins by filtering secondary effluent until a preset headloss is developed. Air mixing is then started in the liquid above the bed to resuspend the solids collected on the media surface. After additional headloss buildup, air trapped in the vented underdrain system is forced upward through the bed for a short period. Solids removed by the air are resuspended by the air mixing, and the cycle begins again. After the predetermined number of cycles, the filter is backwashed, ending the run.

Data from Clark County, Ohio, indicate an average filtrate SS concentration of 4.8 mg/l using effluent from a 0.2 mgd contact stabilization plant at 1.2 gpm/sq ft (57).

9.11 Slow Sand Filters

Slow sand filters consist of a layer of sand supported on graded gravel with an underdrain system but no backwash system. The depth of the sand layer ranges up to 42 inches, and the effective size is 0.25 to 0.35 mm with a uniformity coefficient of 2 to 3 (53). Secondary effluent is applied, generally at a rate of about 3 gph/sq ft (8), and the filter is used until the wastewater rises to the top of the filter wall. The filter is then removed from service, drained, permitted to dry and then cleaned by manually removing the filtered solids.

Truesdale and Birkbeck (58) report only 60 percent SS removal for slow sand filters and a cleaning frequency of once or twice per month. Rapid clogging of slow sand filters has been observed (59). Slow sand filters require large land areas and therefore, are not normally employed for large installations. Sand that is lost during cleaning must eventually be replaced.

Filters of the same construction, operated intermittently, have been used as combined physical-biological treatment for secondary effluent polishing. Intermittent operation permits aerobic digestion of solids reducing somewhat the required frequency of maintenance. Area requirements are still quite large, however, and generally limit applications to small plants. The fact that maintenance is only required on an intermittent basis makes this type of filter a viable process for upgrading existing lagoons which cannot meet effluent standards. Further discussion of this application can be found in the U.S. EPA manual, Upgrading Existing Wastewater Treatment Plants and elsewhere (60).

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CHAPTER 10

COST ESTIMATES

10.1 Introduction

The cost curves included in this chapter are based on: 1) actual installations, 2) projections from pilot studies and other literature, and 3) manufacturers' information. In general, larger capacity units show economy in both capital and operating costs. Since the added economy of scale for plants greater than 100 mgd size is small, unit figures for this flow (or the total area required at this flow) can be applied to larger plants. Costs for plants smaller than 1 mgd vary too widely to permit effective use of general curves.

Cost calculations were based on an EPA-STP Index of 175 (July 1972 for U.S. Average). The procedure for adjustment of costs to another cost index is outlined in Section 10.5.

10.2 Curve Content

The curves shown include all equipment and controls necessary for a working unit process. Construction is assumed to include excavation and backfill in good soil on a level site. In general cost curves do not include:

1. Buildings
2. Land
3. Pumping between processes
4. Sludge disposal
5. Yard piping
6. Special site conditions requiring pile foundations, rock excavation, etc.
7. Chemical feed equipment (given as separate curve)
8. Automated control (except as noted)
9. Engineering, Legal and Fiscal Costs.

10.3 Operation and Maintenance Costs

These are presented as curves or as a percentage of total capital costs and include normal repairs expected during operation of units, but not breakdowns resulting from misapplication or overloading. Chemical costs are not included in operating costs of chemical feed systems, and should be allowed for separately based on actual cost and dosage. O&M costs include power for normal operation but do not include external power costs for pumping between units.

EPA Regulations, Title 40, Chapter 1, Part 35, Appendix A (*Federal Register*, 38, No. 174, September, 1973), specify the useful life of various structures and equipment items to be applied in cost-effectiveness analysis. The regulations also specify use of 7 percent annual interest in cost comparisons. In general, structural items have lives of 30 to 50 years, process equipment 15 to 30 years, auxiliary equipment 10 to 15 years and electrical equipment 8 to 10 years. More specific information than given in the regulations may, in some cases, be obtained from product manufacturers.

10.4 Freight

Costs include freight allowances for equipment based on typical distances in the mid-west and eastern parts of the country. For western installations costs may be up to 2 percent higher due to extra shipping costs.

10.5 How to Use Cost Curves

1. Select capacity of unit(s) based on plant design capacity plus desired standby capacity in unit(s).

2. Enter curve with capacity of unit(s) and read total cost.

3. Correct unit cost to current local cost index. EPA publishes current treatment plant construction cost indices for 20 cities in the U.S. The current cost index should be corrected for time and geographic location. After selecting the projected cost index use the following equation to compute corrected costs:

$$\underset{(\$1000)}{\text{Corrected Total Cost}} = \left[\frac{\text{Current Cost Index}}{175}\right] \times \left[\begin{matrix}\text{Total Cost}\\ (\$1000)\end{matrix}\right]$$

4. The costs arrived at in Step 3 should be modified for the following items wherever applicable:

   a. Special instrumentation for automation or computer control

   b. Special site preparation such as pile foundation, rock excavation, housing and landscaping

   c. Architectural requirements

10.6 Curve Description

The curves are primarily intended for preliminary cost comparisons between processes at any flow or equipment size in the range covered. For actual designs a detailed specific cost analysis should be made. Further data on costs for suspended solids removal processes are available (1) (2) (3) (4) (5) (6).

10.6.1 Flocculators Flash Mixers

Figure 10-1 presents capital and O&M cost estimates for flocculators and flash mixers.

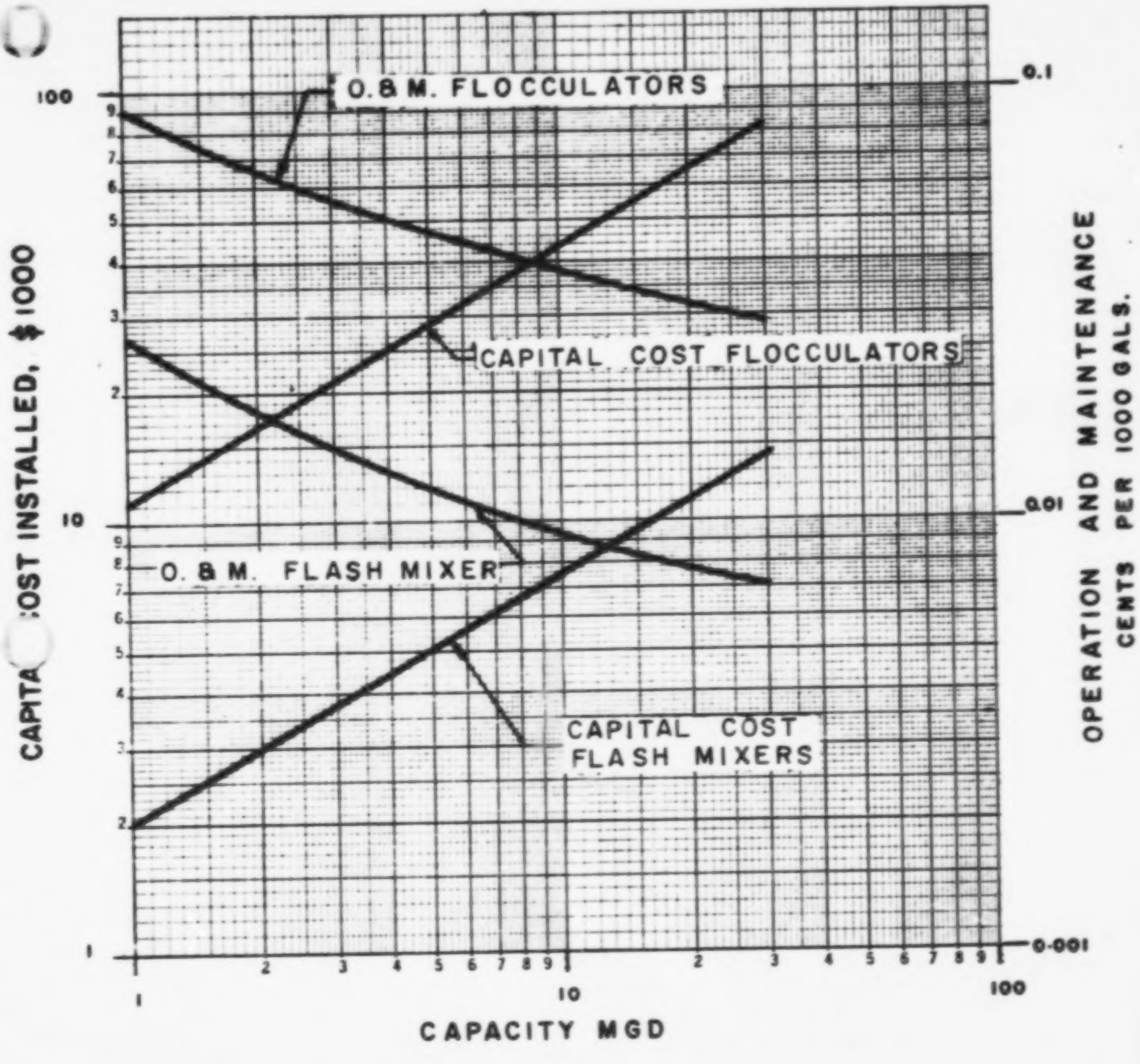

FIGURE 10-1
FLOCCULATORS _FLASH MIXERS
COSTS ADJUSTED TO EPA – STP INDEX 175

Capital costs ($1000) and O&M costs (¢/1000 gal) are given for units in plants of 1 to 30 mgd capacity. Capital costs of both types of units include concrete tanks, floor slabs and walls. Not included in the capital costs are buildings or covers, land, chemical-feed equipment, pumps, and external piping. No instrumentation is included in the estimated costs.

Flash mixer costs are for fully installed vertical shaft, turbine units with stainless steel shafts, stainless steel blades or paddles, gear motors, walkways and supports.

Flocculator costs cover fully-installed horizontal or vertical shaft units with circular or rectangular tank configurations and side water depths of 8 to 12 ft. No special corrosion protection is included except for normally painted items. Flocculator equipment includes gear motors, walkways and supports.

Prices are based on single unit installations; for duplicate installations, use cost for two units, each sized for 50 percent or more of total flow depending on desired standby capacity.

O&M costs for flash mixers were assumed to be 5 percent of the capital costs per year. O&M costs for flocculators were assumed to be 3 percent of the capital costs per year. O&M costs include manpower for operation and normal maintenance but do not include chemicals, power, etc.

10.6.2 Chemical Feed System

Figure 10-2 presents estimates of capital and O&M costs for chemical feed systems.

Capital costs ($1000) are given for alum or iron, lime, and polymer feed systems in plants of 1 to 100 mgd capacity. O&M costs (¢/1000 gal) are given for alum, iron, and polymer feed systems. O&M costs for lime feed systems (not shown) have been reported to be greater than O&M costs for feeding the other chemicals (6) (7) (8). The systems were sized to feed typical rates applied for phosphorus removal.

Dosages, weight basis and stored form of chemicals were assumed as follows:

Chemical	Dosage mg/l	Weight Basis	Storage Form
Alum	200	$Al_2(SO_4)_3 \cdot 14\,H_2O$	50% soln (on wt. basis)
Iron Salts	150	$FeCl_3 \cdot 6\,H_2O$	25% Soln (on wt. basis)
Lime	300	$Ca(OH)_2$	Dry Hydrated Lime $Ca(OH)_2$
Polymer	1	Dry material	Dry material

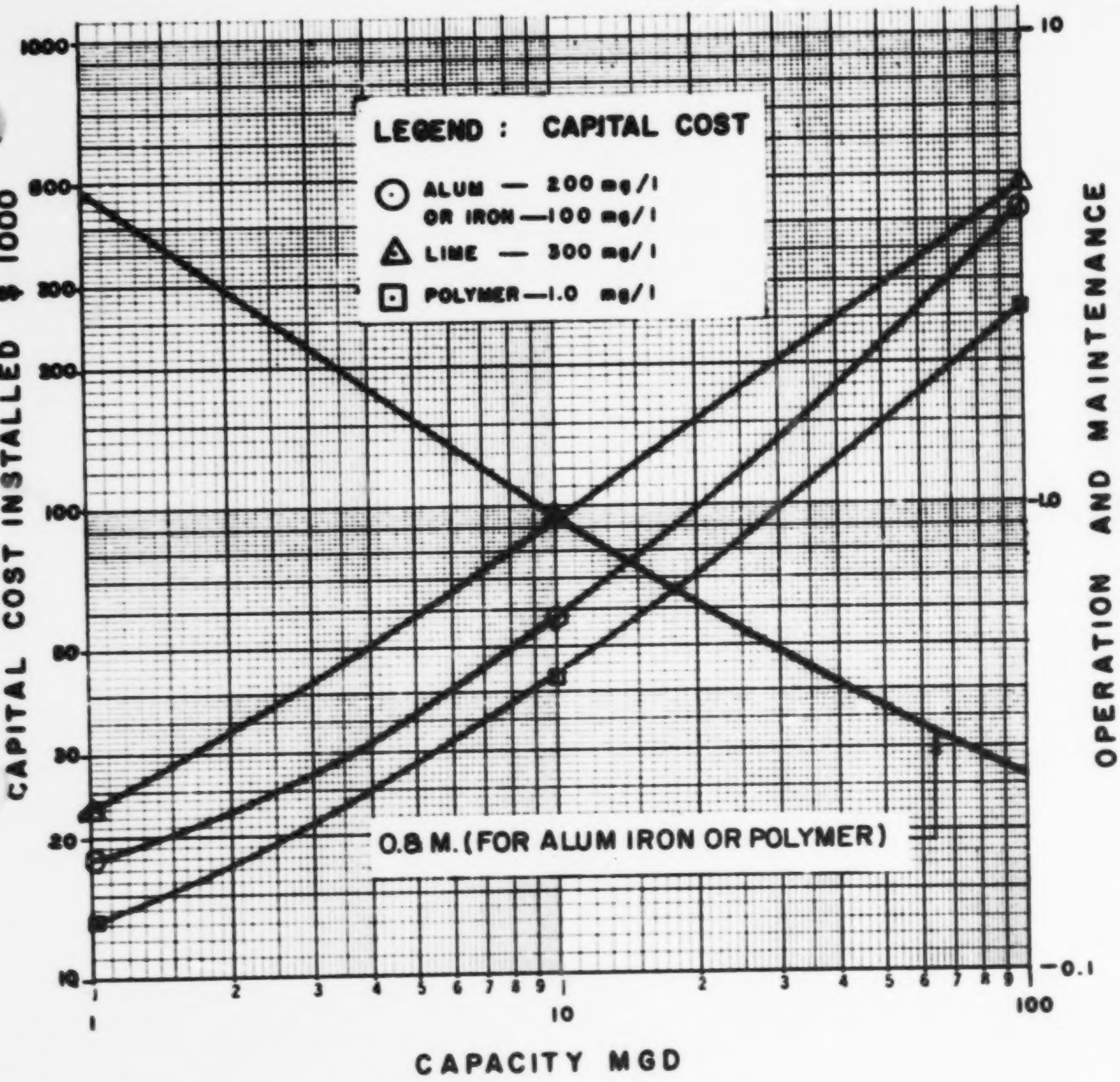

FIGURE 10-2
CHEMICAL FEED SYSTEMS
COSTS ADJUSTED TO EPA STP INDEX 175

For different chemical dosages calculate cost based on an equivalent plant flow calculated as follows:

$$\text{equivalent plant flow} = \left[\begin{array}{c}\text{design}\\ \text{flow}\end{array}\right] \times \left[\frac{\text{expected dosage}}{\begin{array}{c}\text{dosage used to}\\ \text{develop cost curve}\end{array}}\right]$$

Each system includes: 1) a minimum of two volumetric or gravimetric automatic proportioning feeders sized to provide 50 percent excess feed capacity; 2) pumps to deliver chemical feed solutions to the process; and 3) 30-days bulk storage. Prices cover installed equipment suitably corrosion-protected for the intended chemical service. Not included in capital costs are buildings (except for dry chemical storage space), land, sludge disposal and external piping.

O&M costs include manpower for operation and normal maintenance, and power costs for pumping, but do not include chemical costs.

10.6.3 Sedimentation Basins

Figure 10-3 presents estimates of capital and O&M costs for sedimentation basins.

Capital costs ($1000) and O&M costs ($1000/yr) are given for installations requiring total basin surface areas of 1000 to 100,000 sq ft. Costs are based upon installations using two or more units.

Included in capital costs are inlet appurtenances and sludge-collecting mechanisms circular or rectangular tanks (steel or concrete), skimmers, scrapers, supports and walkways, and sludge draw-off, all completely installed. Curves are applicable for end-feed, center-feed or peripheral-feed designs for primary or secondary treatment applications.

O&M costs include manpower for operation and normal maintenance but do not include chemical or pumping costs.

Not included in capital costs are land, buildings, covers, chemical feed equipment, pumping, sludge pumping and disposal and external yard piping. Instrumentation is limited to automatic sludge blowdown valves and lines.

10.6.4 Solids Contact

Figure 10-4 presents estimates of capital and O&M costs for solids-contact units.

Capital costs ($1000) and O&M costs ($1000/yr) are given for installations requiring total basin surface area of 1000 to 100,000 sq ft. A minimum of two operating units was used to estimate costs.

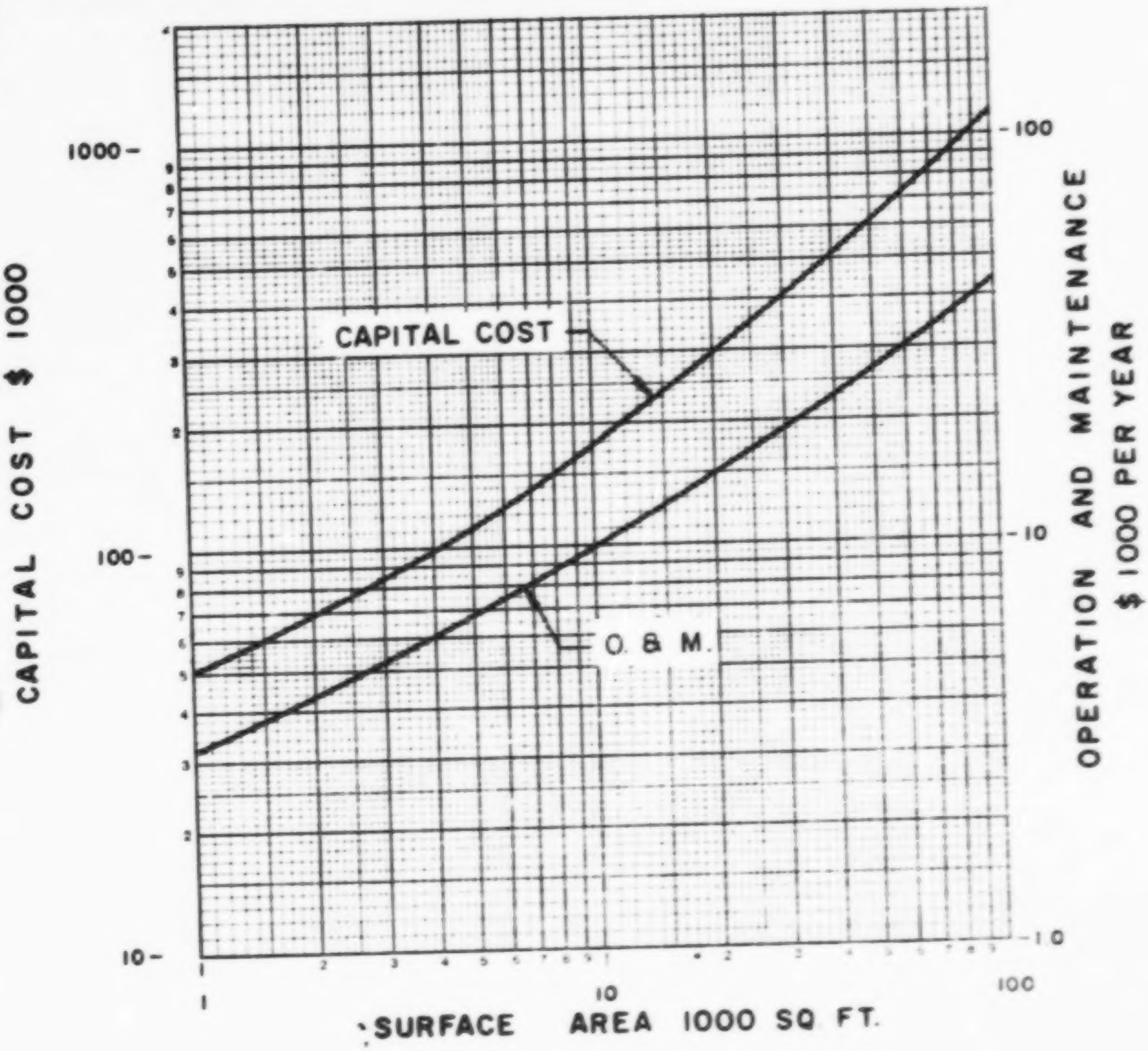

FIGURE 10-3
SEDIMENTATION
COSTS ADJUSTED TO EPA STP INDEX 178

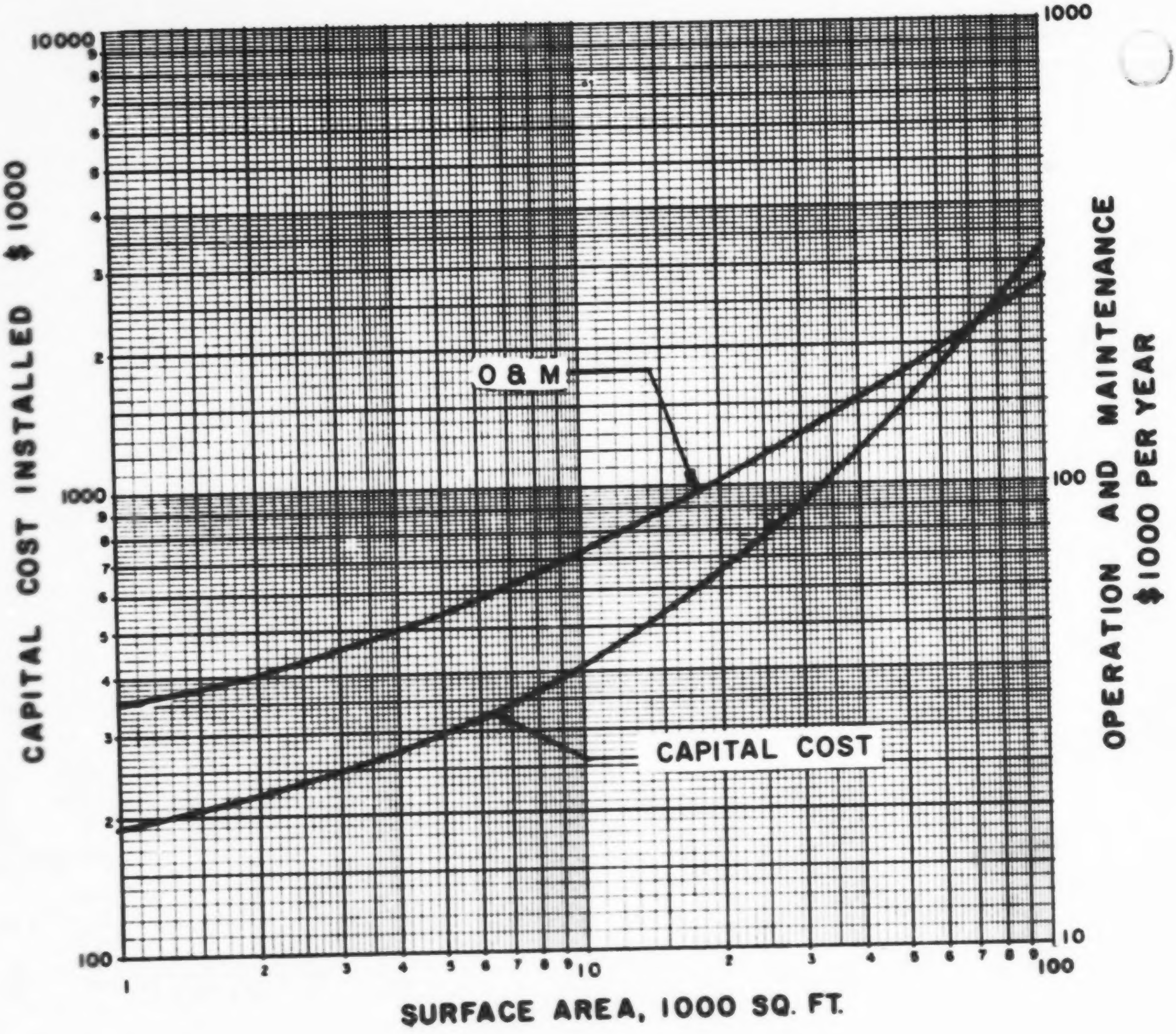

FIGURE 10-4
SOLIDS CONTACT
COSTS ADJUSTED TO EPA—STP INDEX 175

Included in the capital costs are concrete or steel tanks and slabs, turbine recirculators, sludge scrapers, skimmers, inlet and outlet distributors, supports and walkways, sludge drawoff, internal baffles, piping and accessories, all fully installed. Prices are applicable for upflow-type solids-contact units with integral flash mixing and flocculating provisions.

O&M costs include manpower for operation, turbine mixer power, and normal maintenance but do not include chemical or pumping power costs.

Not included in capital costs are buildings, land, chemical-feed equipment, external pumping, sludge disposal and external yard piping. No flow, turbidity, conductivity, or other associated instrumentation is included.

10.6.5 Flotation

Figure 10-5 presents estimates of capital and O&M costs for dissolved-air flotation processes.

Capital costs ($1000) and O&M costs ($1000/yr) are given for installations requiring total basin surface areas of 300 to 50,000 sq ft.

Included in the capital cost are all tanks and internals, air-pressurizing equipment, recycle-pumping equipment, operating valves and piping, all fully installed. No special corrosion protection is included except for normally-painted items.

Assumed O&M costs were 3 percent of capital costs including manpower for operation and normal maintenance, power for normal pumping and air pressurization, but not including chemicals.

Not included in capital costs are buildings, land, chemical feed equipment, sludge disposal and external yard piping. Instrumentation is limited to pressure-sensing controls for normal operation of the units.

10.6.6 Settling Tubes and Wire Septums

Figure 10-6 presents estimates of capital costs for tube settlers and wire septums.

Capital costs are given as $1000 for total required screen areas of 10 to 2000 sq ft. The estimated operating labor requirement is 0.5 man-hrs/day/mgd.

O&M costs for tube settlers can be estimated at 2 man-hr per basin per week. O&M costs for wire septums can be estimated at 1 man-hr per basin per day.

Included in the capital costs of tube settlers are plastic tubes with 60° inclination and 21 in deep plus steel supports and additional effluent collector weirs. Wire septum costs include stainless steel wires, all fully installed.

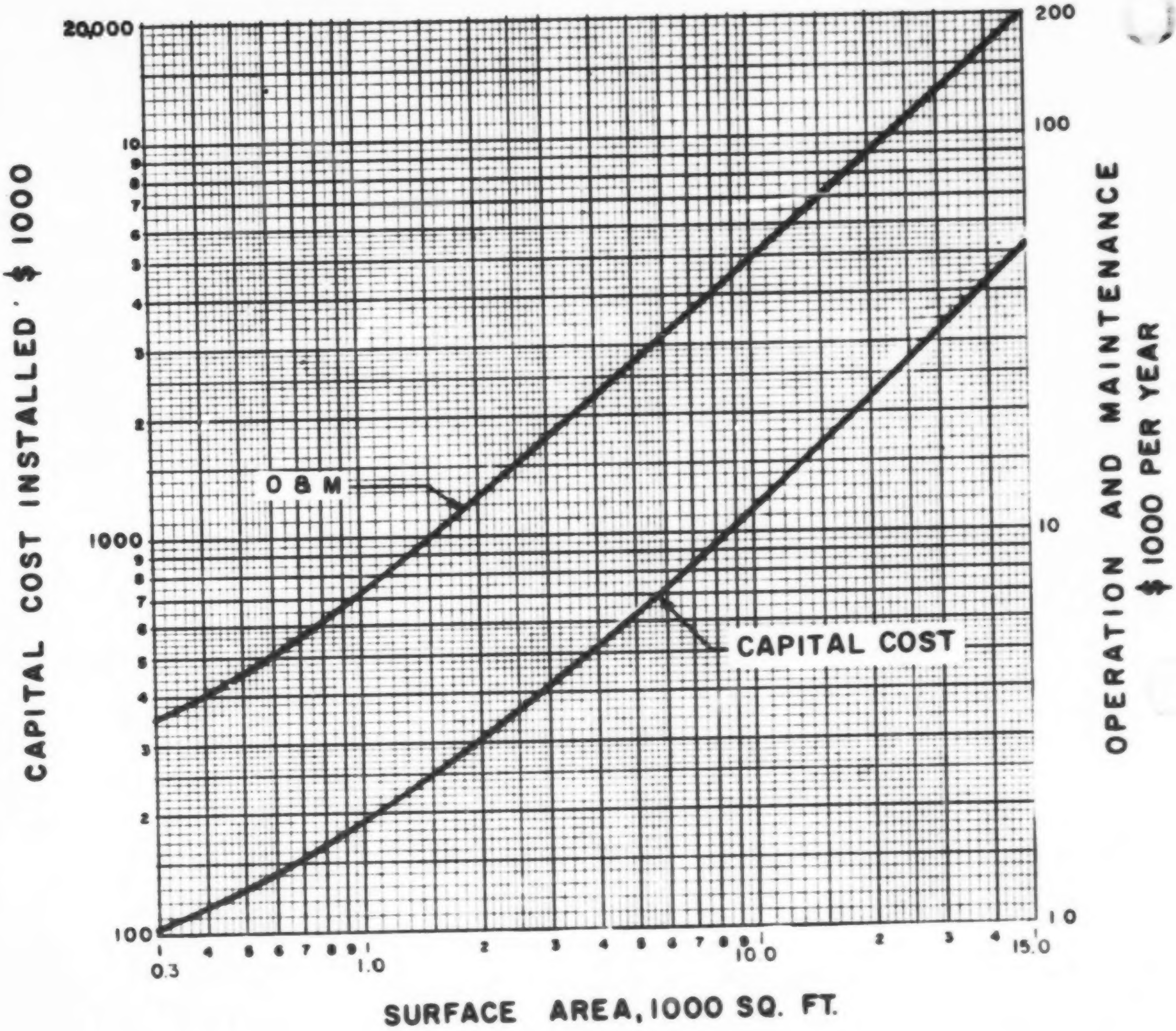

FIGURE 10-5
FLOTATION
COSTS ADJUSTED TO EPA STP INDEX 175

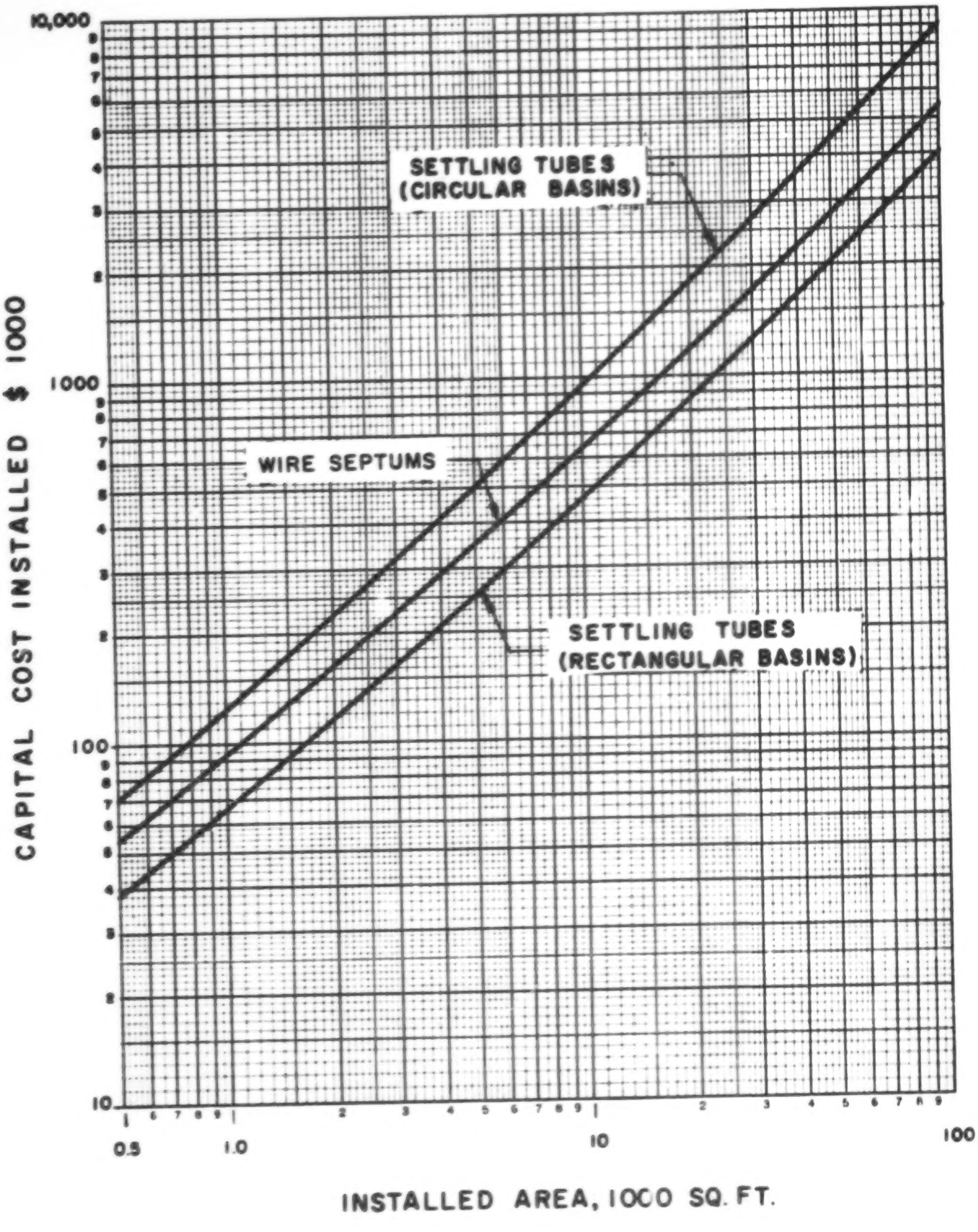

FIGURE 10-6
WIRE SEPTUMS AND SETTLING TUBES
COSTS ADJUSTED TO EPA STP INDEX 175

Prices for wire septums are applicable to circular or rectangular designs. Prices for tube settlers are given separately for circular and rectangular tank designs. No special corrosion protection is included except for normally-painted items.

Not included in capital costs are buildings, tanks, cleaning devices (air grids), sludge disposal and external yard piping.

10.6.7 Wedge-Wire Screens

Figure 10-7 presents estimates of capital costs for wedge-wire screens. O&M costs are calculated as described below.

Capital costs are given as $1000 for total required screen areas of 10 to 2000 sq ft. The estimated operating labor requirements is 0.5 man-hrs/day/mgd.

Included in the capital cost are screens and screen supports. All screens are stainless steel construction, but supports, baffles and distributors are steel. Rotating screens include motor drives. Prices are applicable for rotating screens and stationary screens. No special corrosion protection is included except for normally-painted items.

Not included in capital costs are land, buildings, pumping, sludge handling and external yard piping.

10.6.8 Microscreens

Figure 10-8 presents estimates of capital and O&M costs for microscreen equipment.

Capital costs ($1000) and O&M costs ($1000/yr) are given for installations requiring total screen areas of 100 to 10,000 sq ft. One unit for 1 to 2 mgd flow, two units for 3 to 4 mgd, and three units for flows of more than 5 mgd were used to estimate costs.

Included in the capital costs are tanks, drums, screens, backwash equipment, drive motors and all accessories for automatic operation, all fully installed. Prices are applicable for concrete or steel tank construction. No special corrosion protection is included except for normally-painted items.

O&M costs include operation, normal maintenance, and power for rotation and for spray-water pumping, but do not include chemicals or external pumping power costs.

Not included in capital costs are buildings, land, pumping (except spray system), sludge disposal and external yard piping. Instrumentation is limited to automatic valves and time cycle or pressure sensing backwash control suitably panel mounted.

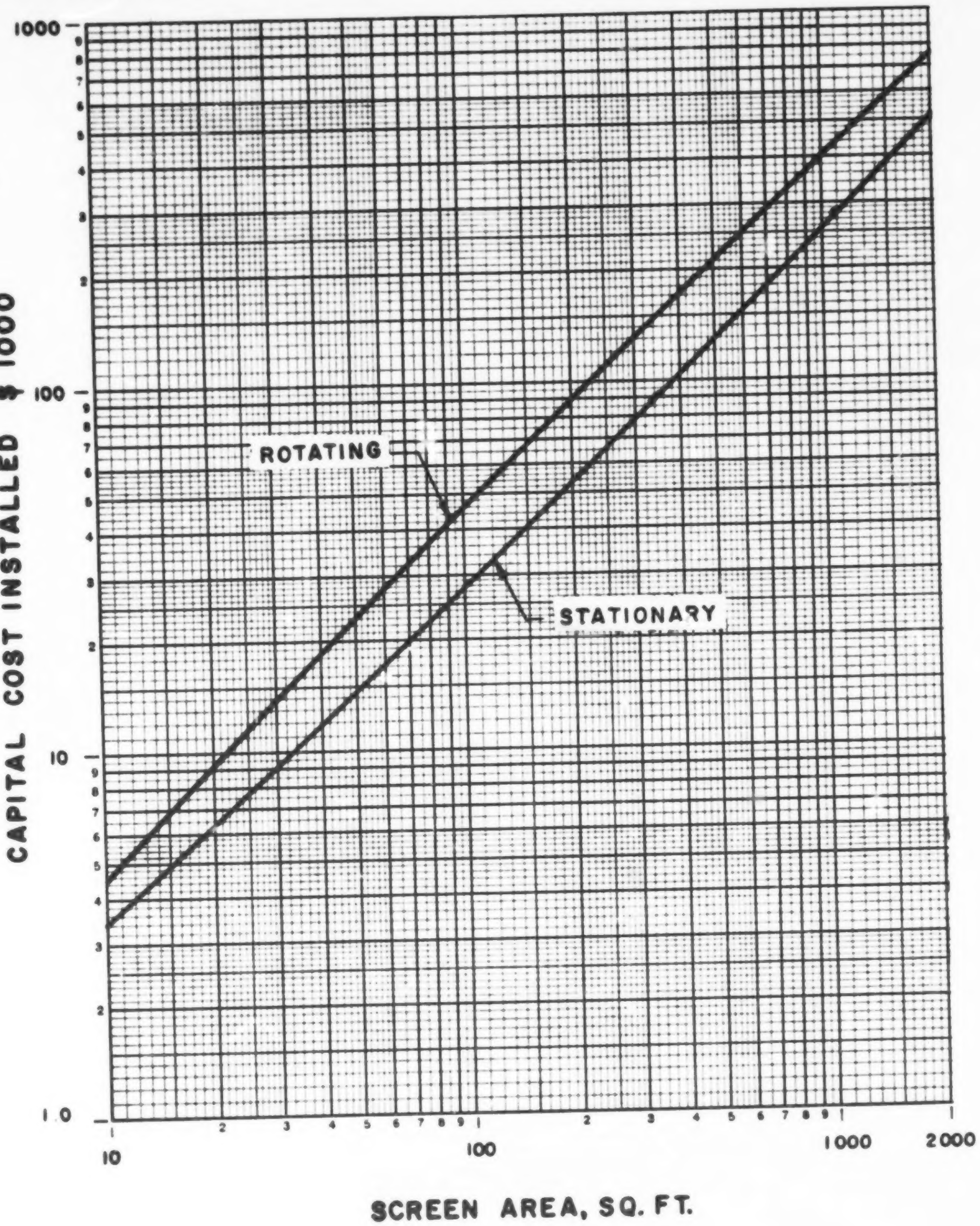

FIGURE 10-7
WEDGE WIRE SCREENS: ROTATING AND STATIONARY
COSTS ADJUSTED TO EPA STP INDEX 175

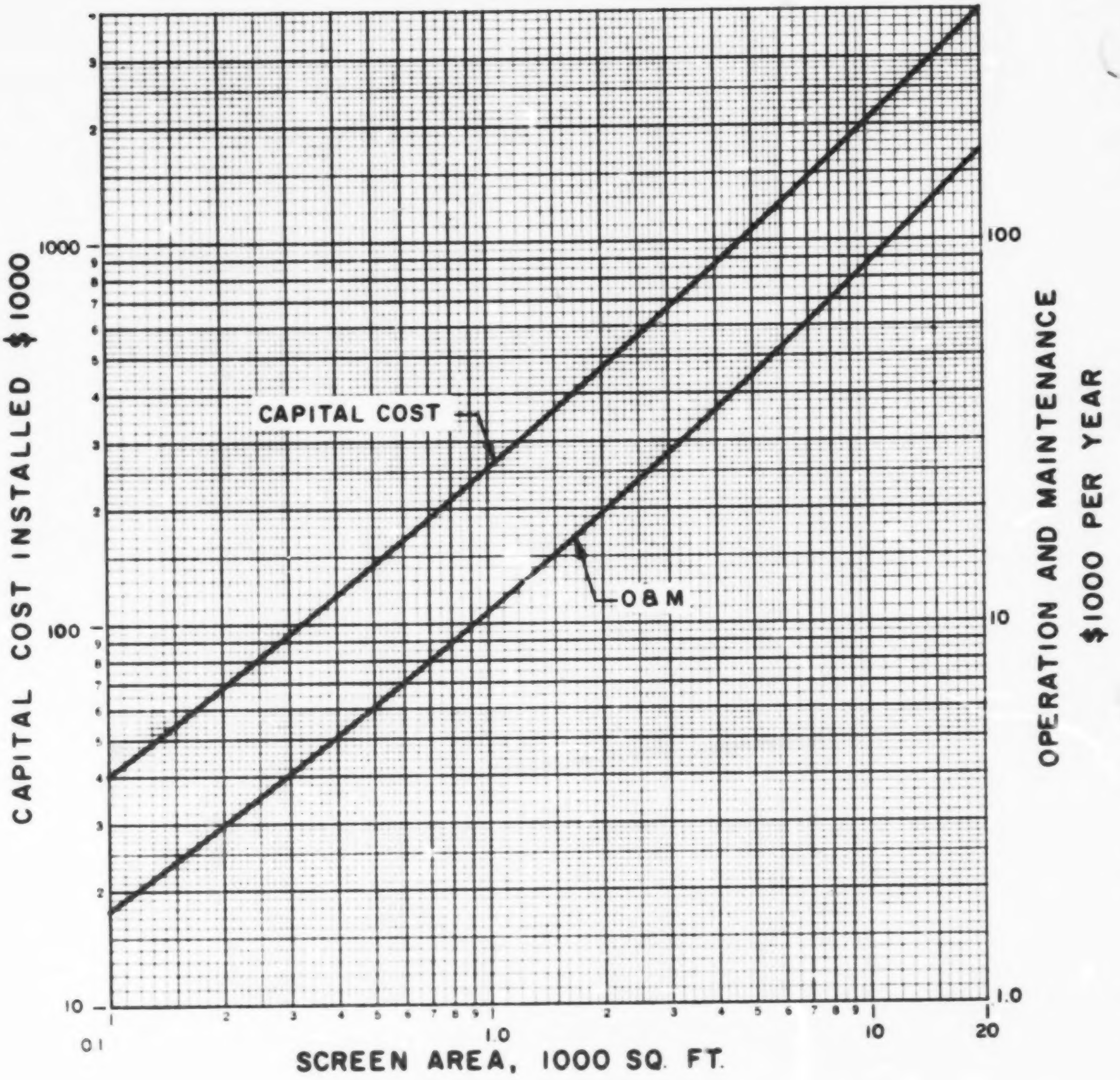

FIGURE 10-8
MICROSCREENS
COSTS ADJUSTED TO EPA STP INDEX 175

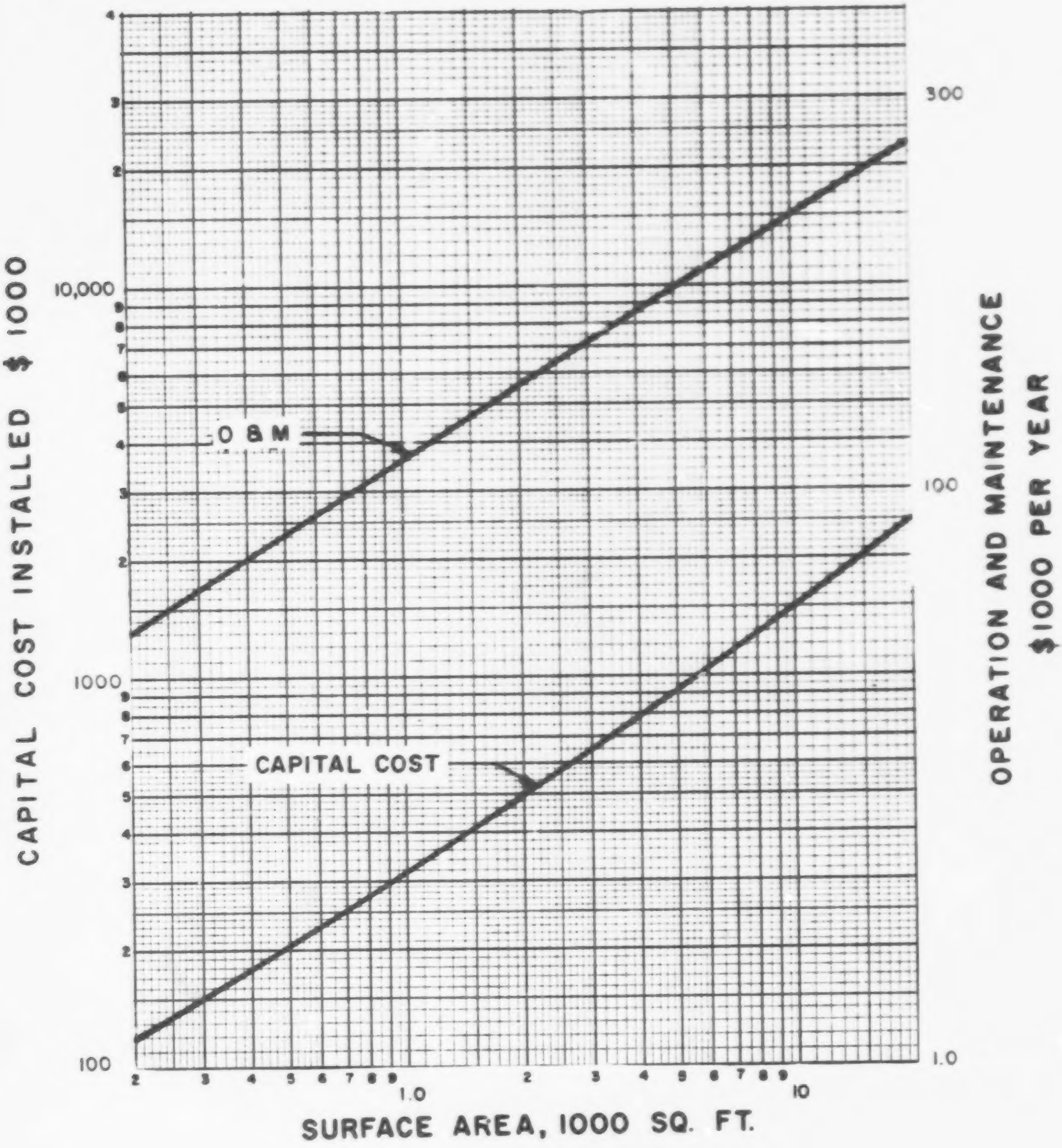

FIGURE 10-9
MEDIA FILTERS
COSTS ADJUSTED TO EPA—STP INDEX 175

10.6.9 Media Filters

Figure 10-9 presents estimates of capital and **O&M** costs for filtration equipment.

Capital costs ($1000) and **O&M** costs ($1000/yr) are given for filter installations with total required surface areas of 200 to 10,000 sq ft. A minimum of three operating units and filter bed depths of 4 to 6 ft. with sand and/or coal media were used to estimate costs.

Included in the capital cost are filter tanks, internals, media, operating valves and piping and automatic backwash controls, all fully installed. The curve shown is an average curve for upflow or downflow, gravity or pressure (up to 60 psig) designs of either concrete or steel construction. Pressure filters are usually less expensive than gravity units below 5 to 6 mgd, but are considerably more expensive at larger flows.

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